U.S. DEPARTMENT OF COMMERCE
National Technical Information Service
PB80-146715

Full-Scale Dual-Alkali Demonstration System at Louisville Gas and Electric Co Final Design and System Cost

Louisville Gas and Electric Co, KY

Prepared for

Industrial Environmental Research Lab, Research Triangle Park, NC

Sep 79

PB80-146715

United States
Environmental Protection
Agency

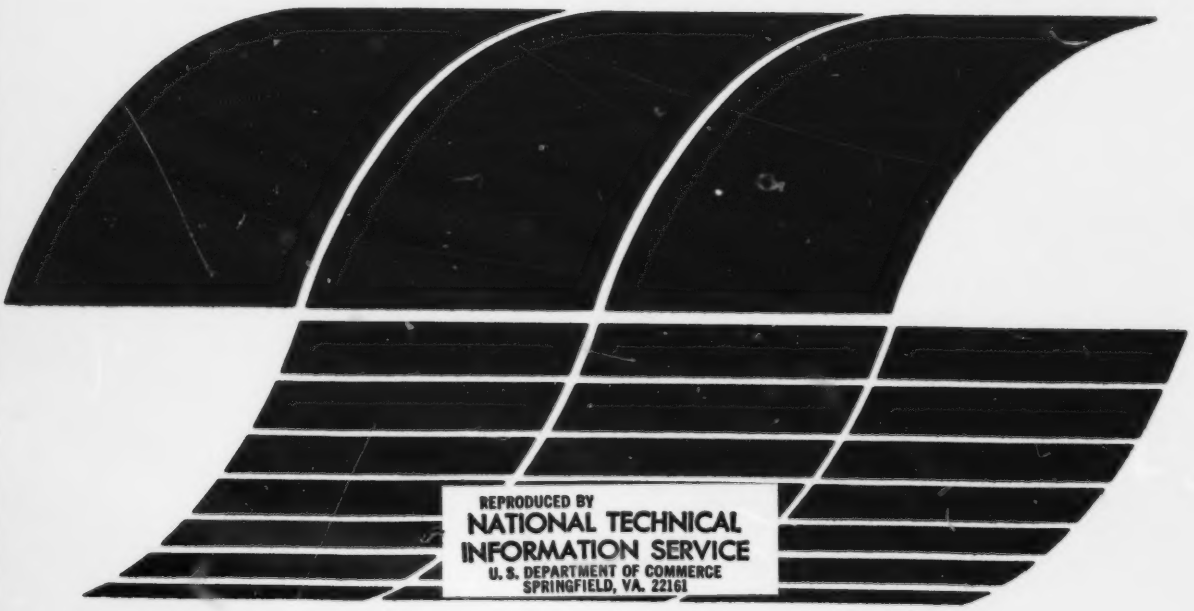
Industrial Environmental Research
Laboratory
Research Triangle Park NC 27711

EPA-600/7-79-221b
September 1979

EPA

Full-Scale Dual-Alkali Demonstration System at Louisville Gas and Electric Co. – Final Design and System Cost

Interagency Energy/Environment R&D Program Report



REPRODUCED BY
NATIONAL TECHNICAL
INFORMATION SERVICE
U. S. DEPARTMENT OF COMMERCE
SPRINGFIELD, VA. 22161

RESEARCH REPORTING SERIES

Research reports of the Office of Research and Development, U.S. Environmental Protection Agency, have been grouped into nine series. These nine broad categories were established to facilitate further development and application of environmental technology. Elimination of traditional grouping was consciously planned to foster technology transfer and a maximum interface in related fields. The nine series are:

1. Environmental Health Effects Research
2. Environmental Protection Technology
3. Ecological Research
4. Environmental Monitoring
5. Socioeconomic Environmental Studies
6. Scientific and Technical Assessment Reports (STAR)
7. Interagency Energy-Environment Research and Development
8. "Special" Reports
9. Miscellaneous Reports

This report has been assigned to the INTERAGENCY ENERGY-ENVIRONMENT RESEARCH AND DEVELOPMENT series. Reports in this series result from the effort funded under the 17-agency Federal Energy/Environment Research and Development Program. These studies relate to EPA's mission to protect the public health and welfare from adverse effects of pollutants associated with energy systems. The goal of the Program is to assure the rapid development of domestic energy supplies in an environmentally-compatible manner by providing the necessary environmental data and control technology. Investigations include analyses of the transport of energy-related pollutants and their health and ecological effects; assessments of, and development of, control technologies for energy systems; and integrated assessments of a wide range of energy-related environmental issues.

EPA REVIEW NOTICE

This report has been reviewed by the participating Federal Agencies, and approved for publication. Approval does not signify that the contents necessarily reflect the views and policies of the Government, nor does mention of trade names or commercial products constitute endorsement or recommendation for use.

This document is available to the public through the National Technical Information Service, Springfield, Virginia 22161.

TECHNICAL REPORT DATA
(Please read Instructions on the reverse before completing)

1. REPORT NO.	2.	3. RECIPIENT'S ACCESSION NO.
EPA-600/7-79-221b		PB 80 146715

4. TITLE AND SUBTITLE	5. REPORT DATE
Full-Scale Dual Alkali Demonstration System at Louisville Gas and Electric Co. -- Final Design and System Cost	September 1979
	6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S) R.P. Van Ness, R.C. Somers, R.C. Weeks, *T. Frank, *G.J. Ramans, **C.R. La Mantia, **R.R. Lunt, and **J.A. Valencia	8. PERFORMING ORGANIZATION REPORT NO.
9. PERFORMING ORGANIZATION NAME AND ADDRESS Louisville Gas and Electric Company 311 W. Chestnut St. Louisville, KY 40201	10. PROGRAM ELEMENT NO. EHE 624 A 11. CONTRACT/GRANT NO. 68-02-2189
12. SPONSORING AGENCY NAME AND ADDRESS **EPA, Office of Research and Development** **Industrial Environmental Research Laboratory** **Research Triangle Park, NC 27711**	13. TYPE OF REPORT AND PERIOD COVERED Final; 9/76 - 3/79 14. SPONSORING AGENCY CODE **EPA/600/13**

15. SUPPLEMENTARY NOTES IERL-RTP project officer is Norman Kaplan, MD-61, 919/541-2556. (*) CEA. (**) A.D. Little, Inc. EPA-600/7-78-010 and -010a are related reports.

16. ABSTRACT

The report describes phase 2 of a 4-phase demonstration program involving the dual alkali process for controlling SO2 emmissions from Unit 6, a coal-fired boiler at Louisville Gas and Electric Co.'s Cane Run Station. The process was developed by Combustion Equipment Associates, Inc., and Arthur D. Little, Inc. The program consists of four phases: (1) preliminary design and cost estimation; (2) engineering design, construction, and mechanical testing; (3) startup and performance testing; and (4) 1-year operation and test programs. The report describes final engineering design, construction and mechanical testing, and installed system capital cost. Construction of the system was completed in February 1979 and system startup was initiated in March 1979. Total capital investment for the entire plant, including waste disposal, is estimated to be $20.4 million (construction of the waste disposal facilities is not complete).

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Pollution Scrubbers Alkalies Sulfur Dioxide Coal Combustion Desulfurization Boilers Design Construction Testing Capitalized Costs	Pollution Control Stationary Sources Dual Alkali Process	13B 14B 13M 07A 07D 14A 07B 21D 21B 13A

18. DISTRIBUTION STATEMENT	19. SECURITY CLASS *(This Report)*	21. NO. OF PAGES
Release to Public	**Unclassified**	
	20. SECURITY CLASS *(This page)* **Unclassified**	22. PRICE

EPA Form 2220-1 (9-73)

EPA-600/7-79-221b

September 1979

Full-Scale Dual - Alkali Demonstration System at Louisville Gas and Electric Co. — Final Design and System Cost

by

R.P. VanNess, R.C. Somers, R.C. Weeks (LG&E);
T. Frank, G.J. Ramans (CEA);
C.R. LaMantia, R.R. Lunt, and J.A. Valencia (ADL)

Louisville Gas and Electric Company
311 W. Chestnut St.
Louisville, KY 40201

Contract No. 68-02-2189
Program Element No. EHE624A

EPA Project Officer: Norman Kaplan

Industrial Environmental Research Laboratory
Office of Environmental Engineering and Technology
Research Triangle Park, NC 27711

Prepared for

U.S. ENVIRONMENTAL PROTECTION AGENCY
Office of Research and Development
Washington, DC 20460

ABSTRACT

The dual alkali process developed by Combustion Equipment Associates, Inc. (CEA) and Arthur D. Little, Inc. (ADL) has been installed in Unit No. 6, a coal-fired boiler at Louisville Gas and Electric Company (LG&E) Cane Run Station for controlling SO_2 emissions. The Federal Environmental Protection Agency (EPA) has selected this system as the demonstration plant for dual alkali technology and is participating in funding of the operation, testing, and reporting of the project.

The project consists of four phases: Phase I - preliminary design and cost estimation; Phase II - engineering design, construction, and mechanical testing; Phase III - startup and performance testing; and Phase IV - one year operation and test programs.

This report covers the work in Phase II of the program including: the final engineering design; construction and mechanical testing; and installed capital cost for the system. Construction of the system was completed in February 1979 and system startup was initiated in March 1979.

CONTENTS

	Page No.
Abstract	iii
List of Figures	vii
List of Tables	viii
Applicable Conversion Factors English to Metric Units	x
Acknowledgements	xi
I. SUMMARY	1
A. Overall Purpose and Scope	1
B. CEA/ADL Dual Alkali System	2
C. Capital Investment	3
II. INTRODUCTION	6
A. Purpose of Project	6
B. Scope of Work	6
C. Project Schedule	8
III. CEA/ADL DUAL ALKALI PROCESS TECHNOLOGY	11
A. Process Chemistry	11
B. Pollution Control Capabilities	17
IV. DESCRIPTION OF THE DUAL ALKALI PROCESS APPLICATION AT THE CANE RUN STATION	19
A. Boiler System Description	19
B. Design Conditions for the Dual Alkali System	19
C. Guarantees	22
D. Process Description	24
E. Material Balances	33
F. Operating and Control Philosophy	41

CONTENTS (Continued)

	Page No.
V. DUAL ALKALI PLANT CONFIGURATION AND EQUIPMENT SPECIFICATIONS	54
A. Plant Layout	54
B. Process Equipment	54
C. Offsites and Auxiliaries	60
D. Mechanical Testing of Equipment	64
VI. CAPITAL COSTS FOR THE DUAL ALKALI SYSTEM AT THE CANE RUN STATION	68
A. FGD System	68
B. Lime Slurry Feed	73
C. Waste Disposal	73
VII. GLOSSARY	74
APPENDIX - EQUIPMENT DETAILS	

FIGURES

Figure No.		Page No.
II-1	Dual Alkali Demonstration Overall Project Schedule	9
II-2	Phase II Schedule	10
III-1	Dual Alkali Process Flow Diagram	13
V-1	Overall View of Cane Run Unit No. 6 and the Dual Alkali System	55
V-2	Gas Reheaters and Absorber Ducting	56
V-3	Overall View of the Chemical Plant	57

DRAWINGS

Process Flow Diagram 040044-1-1, Rev. G

Process Flow Diagram 040044-1-2, Rev. G

TABLES

Table No.		Page No.
I-1	Capital Costs for the Dual Alkali System at LG&E's Cane Run Unit No. 6	5
IV-1	Ultimate Analysis of Coal Fired in Unit No. 6	20
IV-2	Design Basis	21
IV-3	Flue Gas Conditions at the Inlet of the Dual Alkali System	23
IV-4	Carbide Lime Specifications	31
IV-5	Basis for Material Balances at Design Conditions	34
IV-6	Overall Material Balance at Design Conditions	35
IV-7	Overall Water Balance at Design Conditions	36
IV-8	Material Balance - Absorber Section	38
IV-9	Material Balance - Reactor, Solids Dewatering and Raw Materials Section	39
IV-10	Control Philosophy for the Absorber Section	43
IV-11	Control Philosophy for the Reactor Section	46
IV-12	Control Philosophy for the Solids Separation Section	49
IV-13	Control Philosophy for the Raw Materials Section	52
V-1	Ancillary Requirements for the Dual Alkali System at LG&E Cane Run Unit No. 6	62
VI-1	Capital Costs for the Dual Alkali System at Cane Run Unit No. 6	69
VI-2	Capital Cost Breakdown by Sub-System	70
VI-3	Material Costs for the FGD System - Process Materials	71
VI-4	Material Costs for the FGD System - Additional Materials	72

TABLES (Continued)

Table No.		Page No.
A-1	Equipment List	A-1
A-2	Materials of Construction	A-4
A-3	Agitators	A-7
A-4	Dampers	A-8
A-5	Ductwork	A-9
A-6	Expansion Joints	A-10
A-7	Booster Fans	A-11
A-8	Reheaters	A-12
A-9	Pumps	A-13
A-10	Tanks	A-14
A-11	Soda Ash Silo	A-15
A-12	Thickener	A-16
A-13	Vacuum Filter	A-17
A-14	Absorber	A-18

APPLICABLE CONVERSION FACTORS

ENGLISH TO METRIC UNITS

British	Metric
5/9 (°F-32)	°C
1 ft	0.3048 meter
1 ft^2	0.0929 $meters^2$
1 ft^3	0.0283 $meters^3$
1 grain	0.0648 gram
1 in	2.54 centimeters
1 in^2	6.452 $centimeters^2$
1 in^3	16.39 $centimeters^3$
1 lb (avoir.)	0.4536 kilogram
1 ton (long)	1.0160 metric tons
1 ton (short)	0.9072 metric tons
1 gal	3.7853 liters
1 Btu	252 calories

ACKNOWLEDGEMENTS

This report was prepared by Arthur D. Little, Inc.; however, the information and data contained in the report represent the work of many individuals from several organizations who have been involved in this project. The principal participating organizations are Louisville Gas and Electric, Inc., Combustion Equipment Associates, Inc., and Arthur D. Little, Inc.

In addition, we would like to acknowledge the efforts and contributions from persons in other organizations. Norman Kaplan, the EPA Project Officer for this demonstration program, has made important technical contributions and has been instrumental in the management of the entire project. Mike Maxwell, the Director of Emissions/Effluent Technology at EPA's Industrial Environmental Research Laboratory, was responsible for overall planning and review for this program and has provided invaluable guidance and support. And Randall Rush of the Southern Company Services has made important contributions of a technical nature to the design of the system.

I. SUMMARY

A. OVERALL PURPOSE AND SCOPE

The dual alkali process developed by Combustion Equipment Associates, Inc. (CEA) and Arthur D. Little, Inc. (ADL) has been installed in Unit No. 6, a coal-fired boiler at Louisville Gas and Electric Company (LG&E) Cane Run Station for controlling SO_2 emissions. The Federal Environmental Protection Agency (EPA) has selected this system as a demonstration plant for dual alkali technology and is participating in funding of the operation, testing, and reporting of the project.

The project consists of four phases:

I Preliminary Design and Cost Estimation

II Engineering Design, Construction, and Mechanical Testing

III Startup and Performance Testing

IV One Year Operating and Test Program

Construction of the system was completed in February, 1979 and system startup was initiated in March, 1979. At the conclusion of startup, the system will undergo acceptance testing followed by a one year demonstration test program. The test program will be conducted by Bechtel National, Inc. (Becthel) under contract to EPA. The objectives of the demonstration program are:

- Characterize the system operation over the range of conditions encountered on Cane Run Unit No. 6 including exploratory testing of the effect of fly ash inclusion and the substitution of commercial lime for carbide lime on system chemistry;
- Monitor the performance of the system relative to design criteria and performance guarantees set forth in the LG&E EPA contract;
- Evaluate the overall technical and economic feasibility/applicability of the system;
- Demonstrate long-term reliability;
- Investigate environmentally acceptable methods of waste disposal; and
- Study the effect of the system (in a generic sense) on plume dispersion.

This report covers the work in Phase II of the program including: the final engineering design; construction and mechanical testing; and installed capital cost for the system.

B. THE CEA/ADL DUAL ALKALI SYSTEM

1. General System Description

The dual alkali system involves absorption of SO_2 using an aqueous solution of alkaline sodium salts. Regeneration of the spent absorbent solution is accomplished using lime which produces a solid waste of calcium-sulfur salts. The system operates in a closed loop, the only waste material being a washed filter cake--there are no other solid waste or liquid purged streams. In most applications, the system involves basically four process subsections: flue gas contacting (SO_2 absorption); absorbent solution regeneration and formation of waste solids; waste solids dewatering (thickening and filtration); and raw materials storage and feed preparation (for soda ash and lime). In some cases, the waste filter cake may be further processed in a treatment facility to produce a stabilized material for disposal.

The process operates in a concentrated active sodium mode[a] and is capable of SO_2 removal efficiencies in excess of 95% over a range of inlet SO_2 concentrations normally encountered in coal fired utility boiler applications. In addition to SO_2 removal, the system is highly effective in absorption of chlorides from the flue gas; and particulate removal can also be accommodated by appropriate scrubber selection.

2. Application to Cane Run Unit No. 6

Cane Run Unit No. 6 is a high sulfur, coal fired boiler having a gross peak capacity of 300 Mw. The sulfur content of the coal burned in this unit ranges from about 3.0% to 6.3% (dry basis); and the chloride content typically varies from 0.03% to 0.06% (dry basis). The system is designed for SO_2 control only and is capable of operating with coal sulfur contents in excess of 5.0% (dry basis). Design flue gas conditions down stream of the induced draft fan and at the inlet of the dual alkali system are as follows:

- two parallel trains
- gas flow rate--1,065,000 acfm
- temperature--300°F
- pressure--minus 1 to plus 2 inches W.G.
- SO_2 concentration--3,471 ppm (dry basis)
- particulate matter concentration--$\leq$0.0537 grains/acf

[a]See glossary for definition of dual alkali terminology.

The system is modular in nature and includes two absorber modules, two reactor trains (each train consisting of two reactors in series), one thickener, and three filters. The absorption system can be operated from 20% to 100% of the gross peak capacity. At loads less than 60% of the design capacity, the system can be operated with a single absorber module. The system can also be operated with one reactor train at 100% load for short durations under design conditions. Spare capacity in the system includes: pumps--100%; filters--50%; and instrumentation for operation--100%. Normally, the system will utilize locally available carbide lime, a byproduct of acetylene production. However, commercial lime will also be used for some test periods during the demonstration program. Waste filter cake produced by the system will be stabilized in a separate waste processing plant to be installed by IU Conversion Systems, Inc. (IUCS). The stabilized material will then be disposed of in landfills adjacent to the plant.

The CEA/ADL dual alkali system has been designed to meet the following process performance guarantees:

- Emissions from the system shall be no greater than 200 ppm SO_2 (dry basis) at coal sulfur contents of up to 5.0%; and the system will provide 95% SO_2 removal for coal containing greater than 5.0% sulfur.
- The system will cause no increase in particulate matter in the flue gas.
- Consumption of lime will not exceed 1.05 moles of available CaO per mol of SO_2 removed from the flue gas.
- Sodium make up will not exceed 0.045 moles of Na_2CO_3 per mole of SO_2 removed from the flue gas.
- The system will consume less than 1.2% of the power generated by the boiler at peak boiler load.
- The filter cake will contain a minimum of 55% insoluble solids.
- The system will have an availability[b] of at least 90% for a one year period.

C. CAPITAL INVESTMENT

The installation of the dual alkali system at Cane Run Station Unit No. 6 required capital investment in three different facilities: the flue gas desulfurization (FGD) system itself, the lime slurry feed system, and a waste processing and disposal system. Each of these facilities have

[b]Availability is defined as the ratio of the hours the system is available for operation and the total hours in the operating period (expressed as a percentage).

involved independent design and installation efforts, and their costs are reported separately. While construction of the FGD system was essentially completed in February, 1979, a significant amount of work remained on the lime feed and waste processing facilities. Therefore, the costs presented here represent the actual expenditures incurred plus estimates for completion of the system.

Table I-1 gives a summary of the capital investment for all three facilities. The total projected cost of $20.6 million includes actual expenditures reported through February 28, 1979 and the estimated capital required for completion. Approximately 80% of this total projected capital cost was expended through the end of February. Most of the remaining 20% is related to the waste processing plants.

TABLE I-1

CAPITAL COSTS FOR THE DUAL ALKALI SYSTEM AT LG&E'S CANE RUN UNIT NO. 6 ($000)*

Subsystem	Material Costs†	Erection and Engineering Costs	Total
FGD	10,256	6,207	16,463
Lime Slurry Feed	800	416	1,216
Waste Disposal	1,959	959	2,918
			20,597

*As-incurred costs plus estimate for completion basis.

†Including spare parts

II. INTRODUCTION

A. PURPOSE OF PROJECT

The project covers the full scale application of the CEA/ADL dual alkali flue gas desulfurization (FGD) system to Unit No. 6, a coal-fired boiler at Louisville Gas and Electric Company's (LG&E) Cane Run Station in Louisville, Kentucky.

The system has been installed on this existing 300 Mw (gross peak capacity) unit to comply with requirements[a] of the Jefferson County Air Pollution Control District, the Kentucky State Division of Air Pollution, and Region IV of the U.S. EPA. EPA selected the dual alkali SO_2 control process at LG&E as a demonstration system for dual alkali technology and is participating in funding of the operation, testing, and reporting of the project.

The dual alkali system has the capability to control the SO_2 emissions to less than 200 ppm dry basis without additional air dilution when burning coal containing up to 5% sulfur. When burning coal containing greater than 5% sulfur, the system will remove at least 95% of the sulfur dioxide in the inlet flue gas. The dual alkali system is not designed for removal of particulate matter; however, it is designed not to increase the loading of particulate matter in the flue gas. As a demonstration system, the purpose of the installation and operation is to establish:

- overall performance - SO_2 removal, lime utilization, sodium makeup, regeneration of spent liquor, water balance, scaling and solids buildup problems, materials of construction, waste cake properties, reliability, and availability.

- economics - capital investment and operating cost.

B. SCOPE OF WORK

The scope of work for the project includes the design, construction, startup, acceptance testing, and one year of operation of a CEA/ADL concentrated mode dual alkali system on Unit No. 6, a 280 Mw coal-fired boiler at LG&E's Cane Run Station. The system is to be designed to treat all of the flue gas emitted at the nominal rated capacity (280 Mw) with the capability for treating flue gas equivalent to a minimum boiler load of 60 Mw and a maximum load of 300 Mw.

---

[a]Removal of 85% of the SO_2 present in the flue gas at the scrubber inlet.

LG&E is the prime contractor with overall responsibility for all aspects of the project. CEA, as a subcontractor to LG&E, is responsible for the engineering design, for the supply of all process equipment, and for engineering assistance during startup and acceptance testing. CEA is also responsible for compliance with all process guarantees and equipment warranties. ADL, a subcontractor to CEA, will provide process engineering support to CEA during design, startup, and acceptance testing; and will provide process assistance to LG&E in the operation of the system during the one year test program. ADL is also responsible for the preparation of all reports required under the EPA/LG&E contract.

The work is divided into four phases:

- Phase I - preliminary design and cost estimates;
- Phase II - engineering, design, construction, and mechanical testing;
- Phase III - startup and performance testing; and
- Phase IV - one year of operation and testing.

Baseline testing on the boiler and monitoring of the system performance during acceptance testing and the one year test program is not included as a part of this contract. This work will be carried out by Bechtel under a separate contract with EPA.

This report covers work performed in Phase II of the project. During this phase, LG&E/CEA/ADL were to:

- complete all aspects of the detailed engineering including material and equipment specifications;
- fabricate or procure the materials and equipment;
- construct the dual alkali plant;
- provide a system for disposal of all waste products from the plant operation;
- provide all spare parts, maintenance supplies, and operating materials;
- demonstrate the mechanical acceptability of the plant;
- prepare an operating manual; and
- select and train operators to properly operate the demonstration plant and establish and properly staff a control laboratory.

This report describes in detail the plant, as built, and its planned mode of operation. It includes: a description of the process; the operating and control philosophy; material balances and utility requirements; plant layout; a description of major items of process equipment; a description of offsites and auxiliaries; results of the mechanical testing; and actual capital costs for the system.

C. PROJECT SCHEDULE

The overall project schedule covering all phases of the dual alkali demonstration project is given in Figure II-1. The overall project, including the one-year test program, was originally scheduled for 40 months (with an additional one month for completion of the final draft report).

Phases I and II were scheduled to begin simultaneously to expedite the overall project. Phase I (preliminary design) was scheduled for five months including preparation of the draft report. Phase II (engineering design and construction) was scheduled for 24 months starting with the signing of the contract. A detail of the schedule for Phase II is given in Figure II-2.

As indicated in Figures II-1 and II-2, the project has been delayed due to the severe winter of 1977/1978. The projected schedule for completion of the project is shown in Figure II-1.

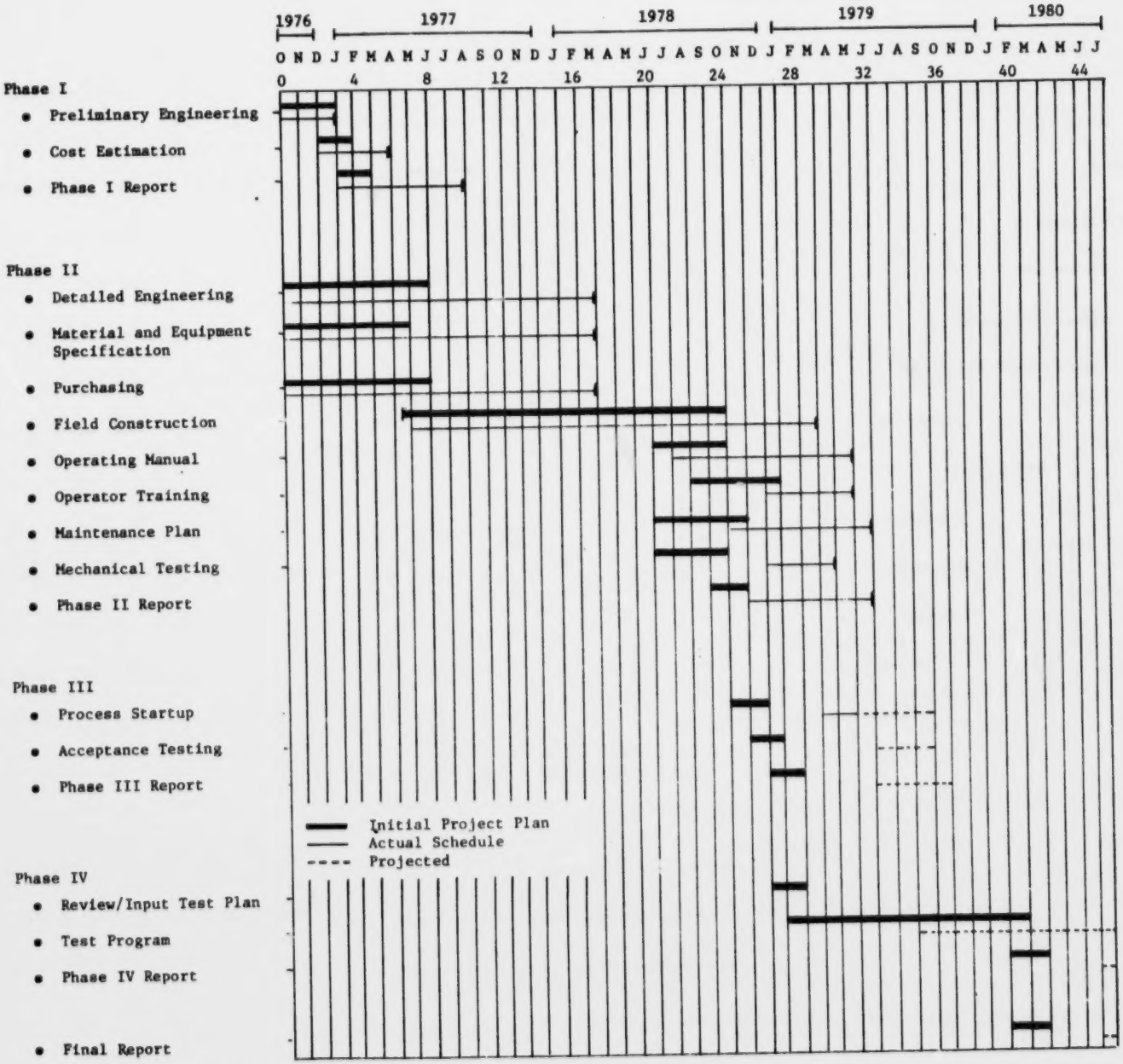


Figure II-1: Dual Alkali Demonstration Overall Project Schedule

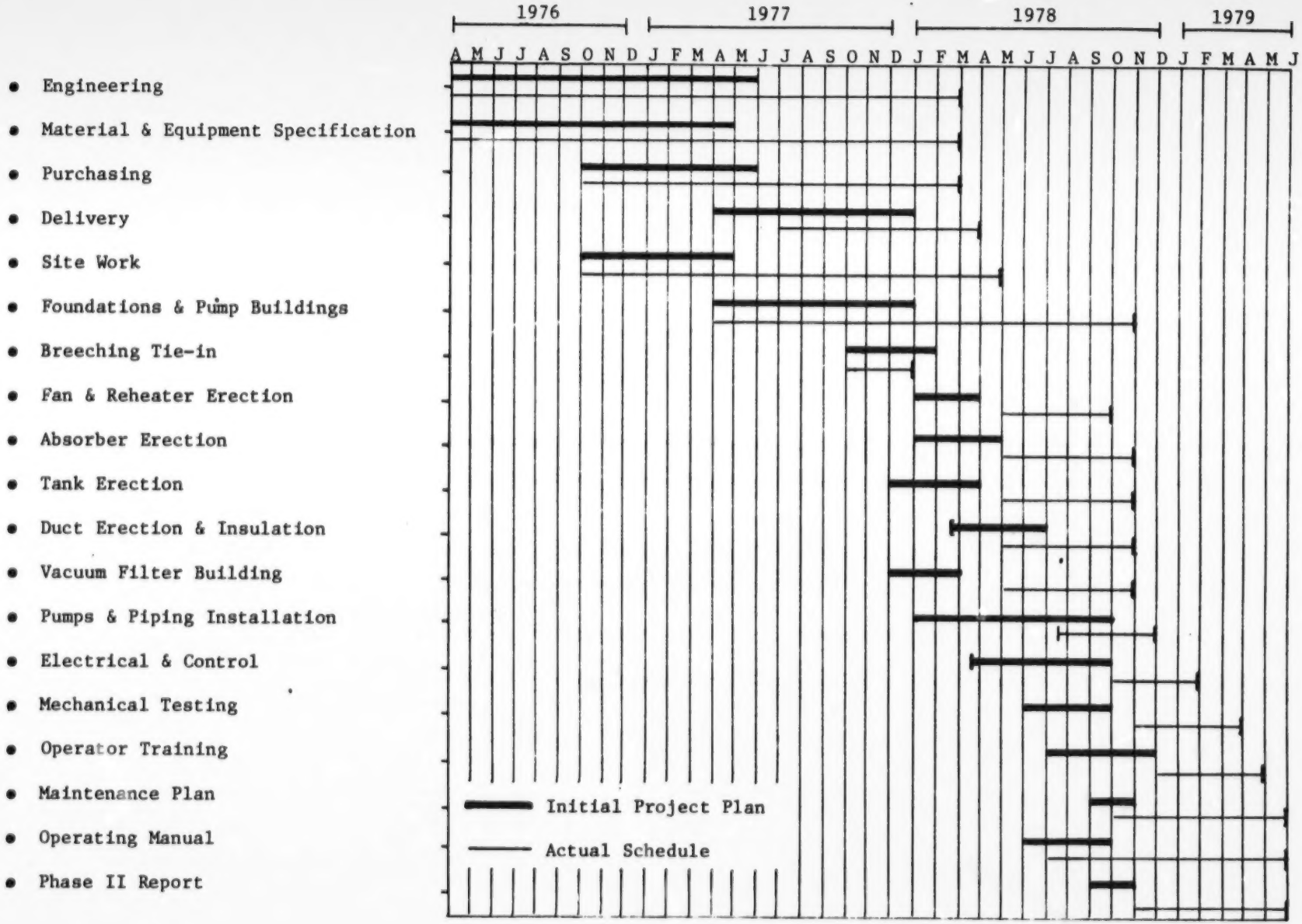


Figure II-2: Phase II Schedule

III. CEA/ADL DUAL ALKALI PROCESS TECHNOLOGY

Dual alkali (or double alkali) is a generic term used to describe flue gas desulfurization (FGD) systems involving the absorption of SO_2 using a soluble alkali, followed by reaction of the spent scrubber solution with lime and/or limestone to regenerate the alkali and produce a waste calcium-sulfur salt for disposal.

The two principal features of dual alkali technology which set it apart from conventional direct lime and limestone scrubbing are: (1) the use of a clear solution rather than a slurry for contacting the flue gas in the absorber; and (2) reaction of the solution in a separate absorbent regeneration section to form the waste solids rather than forming the waste solids as a part of the scrubbing operation. The use of solution (rather than slurry) scrubbing has a number of important advantages. First, with alkaline solutions, high SO_2 removal efficiencies (95%+) can be easily achieved over a wide range of inlet SO_2 concentrations. Second, the precipitation of the waste calcium sulfite/sulfate solids is performed outside the scrubber circuit in a specially designed reactor system. The control of the crystallization reactions allows for the formation of waste solids with good dewatering properties. Finally, since SO_2 absorption is accomplished using clear solutions and precipitation reactions occur outside the scrubber, there is minimal scale potential in the scrubber circuit. Hence, there is no need for washing the mist eliminator to prevent solids deposition and scale formation.

A. PROCESS CHEMISTRY

The dual alkali system installed on Unit No. 6 at Cane Run Station utilizes alkaline solutions of sodium salts for scrubbing the gas and absorbing SO_2. The solution is regenerated using carbide lime, a waste product from the production of acetylene.

The process is designed to operate as a concentrated-mode dual alkali system (in contrast to a dilute mode). The term "concentrated-mode" indicates the range of concentration of alkaline sodium salts in the absorbent liquor. In a concentrated-mode system the concentration of the absorbent solution is such that the precipitation of gypsum ($CaSO_4 \cdot 2H_2O$) as a separate crystalline phase does not occur during absorbent regeneration, so the system operates unsaturated with respect to gypsum.

Based on the major steps of the dual alkali technology, the flue gas desulfurization system can be broken down into four process areas: gas scrubbing; regeneration of scrubbing solution; solids separation; and raw materials preparations. Some variations in equipment and operation of the system would be expected for different applications of the dual alkali technology. The following description represents a

concentrated-mode system utilizing alkaline sodium solutions for the removal of SO_2 with commercial, slaked lime for regeneration of the sodium solution. A generalized process flow diagram is shown in Figure III-1.

1. Flue Gas Scrubbing

Dual alkali systems are, in general, capable of simultaneous particulate matter and SO_2 control as are most wet scrubbing nonrecovery FGD systems. However, Cane Run Unit No. 6 is equipped with an existing high efficiency electrostatic precipitator for control of particulate matter. Therefore, the following description shall be restricted to that of a scrubbing system designed for SO_2 removal only.

In the absorption section of the system, SO_2 is removed from the flue gas by contacting the gas with a solution of sodium salts. This is usually accomplished in a tray tower equipped with quench sprays for cooling and humidifying the gas. The scrubbed gas is then reheated to prevent condensation and corrosion in the ducts and stack and to improve atmospheric dispersion after being exhausted from the stack.

The alkaline solution used to remove SO_2 from the flue gas contains sodium sulfite (Na_2SO_3), hydroxide (NaOH), carbonate (Na_2CO_3), sulfate (Na_2SO_4), and chloride (NaCl). During the process of removing SO_2, the carbonate, hydroxide, and some sulfite are consumed resulting in a spent sodium sulfite/bisulfite liquor. The SO_2 removal process can be represented by the following overall reactions:

$$2Na_2CO_3 + SO_2 + H_2O \rightarrow Na_2SO_3 + 2NaHCO_3 \quad (1)$$

$$NaHCO_3 + SO_2 \rightarrow NaHSO_3 + CO_2\uparrow \quad (2)$$

$$2NaOH + SO_2 \rightarrow Na_2SO_3 + H_2O \quad (3)$$

$$Na_2SO_3 + SO_2 + H_2O \rightarrow 2NaHSO_3 \quad (4)$$

Although the actual reactions within the absorber are more complex, involving various intermediate ionic dissociations, the above set of simplified, overall reactions is an accurate representation of the overall consumption and generation of the various components.

Sodium sulfite plays the most important role in the absorption of SO_2 since it is usually present in the greatest concentration. The hydroxide and carbonate are present in the absorber feed only in small amounts. The concentration of these three alkaline components is a measure of the SO_2 removal capacity of the absorbing liquor. This capacity is conveniently expressed in terms of the "active sodium" concentration where [active sodium] = 2 x $[Na_2SO_3]$ + [NaOH] + 2 x $[Na_2CO_3]$. It must be pointed out that the use of the term "active sodium" is simply one of convenience since it is only an indirect indication of the absorptive capacity of the liquor. SO_2 is actually absorbed by or reacts with the sulfite, hydroxide, or carbonate ions rather than the sodium ion. The dual alkali system for Cane Run Unit No. 6 is designed to operate at an "active sodium" concentration of 0.45 M in the absorber feed.

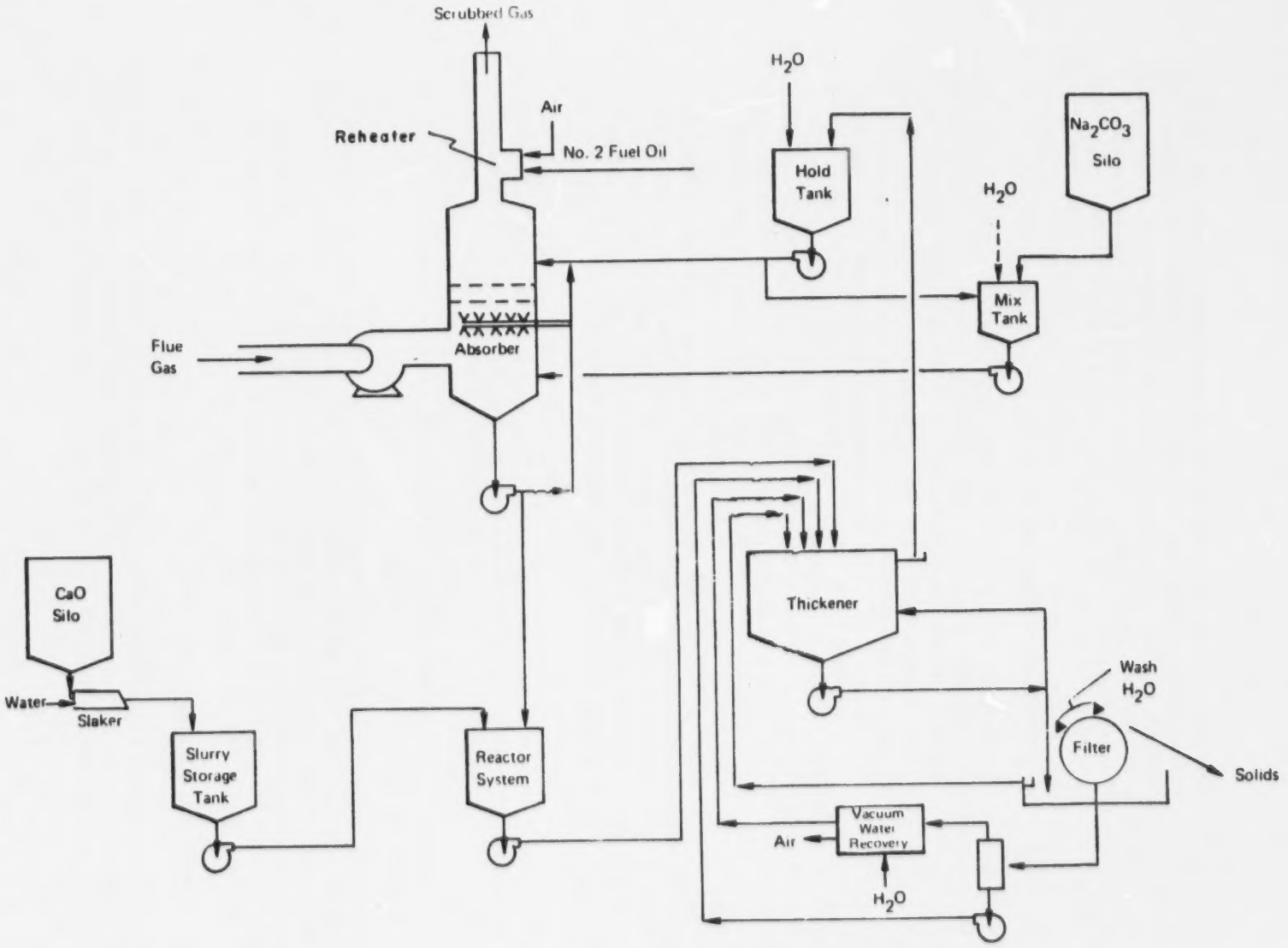


Figure III-1: Dual Alkali Process Flow Diagram

Sodium sulfate and sodium chloride do not participate in the SO_2 removal process. In this sense, they are considered "inactive" components. The presence of sodium sulfate and sodium chloride is principally the result of secondary absorption reactions. Sodium sulfate is formed by the oxidation of sodium sulfite via reaction with oxygen absorbed from the flue gas. Oxidation also occurs in other parts of the system where process solutions are exposed to air; however, the amount of oxidation is small relative to the oxidation which occurs in the absorber.

For a concentrated-mode dual alkali system, the rate of oxidation in the absorber is proportional to oxygen mass transfer, which is a function of the absorber design, oxygen concentration in the gas, gas temperature, and the nature and concentration of the species in the scrubbing solution. For a given set of process parameters, the oxidation rate in moles of sulfite oxidized per unit of time is relatively independent of the SO_2 removal rate. For convenience, though, the amount of oxidation is frequently expressed, on an equivalent basis, as a percentage of the SO_2 removed. For example, in the case of a high sulfur, coal-fired utility boiler with a flue gas containing about 4-5% O_2 and about 2500 ppm SO_2, on the order of 5% to 10% of the SO_2 removed from the flue gas would be expected to be oxidized and appear as sulfate in the spent scrubbing solution. The remaining 90% to 95% would appear in the spent scrubbing solution as sulfite/bisulfite. Under similar conditions of absorber design and solution characteristics, much higher relative oxidation rates can be encountered at higher oxygen concentrations in the gas (higher oxygen mass transfer rates) or at much lower SO_2 removal rates (as in low sulfur coal applications).

At steady state, the sulfate must leave the system either as calcium sulfate or as a purge of sodium sulfate at the rate at which it is being formed in the system. As will be discussed later, relative oxidation rates as high as about 25-30% of the SO_2 removed can be tolerated in concentrated-mode systems without intentional purge of sodium sulfate with the waste solids.

Sodium chloride is formed in the absorber by the reaction of chloride, present in the flue gas as HCl vapor, with the alkaline sodium solutions. The level of sodium chloride in the system builds up to a steady state concentration, such that the rate at which sodium chloride leaves the system with the washed filter cake is equivalent to the rate at which it is picked up in the absorber.

2. Absorbent Regeneration

The SO_2 absorptive capacity of the spent scrubbing solution bleed stream is regenerated in this section of the dual alkali system. The regeneration is accomplished in a two-stage reactor system using carbide lime slurry.

The regeneration of acidic sodium sulfite/sulfate scrubber effluent solutions can be envisioned as a two-step process involving neutralization of bisulfite to sulfite followed by conversion of sulfite to hydroxide:

$$Ca(OH)_2 + 2NaHSO_3 \rightarrow CaSO_3 \cdot 1/2\ H_2O\downarrow + Na_2SO_3 + 3/2\ H_2O \qquad (5)$$

$$Ca(OH)_2 + Na_2SO_3 + 1/2\ H_2O \rightarrow CaSO_3 \cdot 1/2\ H_2O\downarrow + 2NaOH \qquad (6)$$

The first of these reactions is a neutralization reaction which goes to completion. The second is a precipitation reaction in which the equilibrium hydroxide concentration is limited by the relative solubility products for calcium sulfite and calcium hydroxide, and the concentrations of hydroxide and sulfite in solution. Both of these reactions result in the formation of calcium sulfite solids. The usual form of the calcium sulfite is as the hemihydrate salt (calcium sulfite $\cdot 1/2\ H_2O$).

Simultaneously with the neutralization and precipitation reactions indicated above, a limited amount of calcium sulfate will also be precipitated:

$$Ca^{++} + SO_4^{=} + xH_2O \rightarrow CaSO_4 \cdot xH_2O\downarrow \qquad (7)$$

In a concentrated-mode dual alkali system, the sulfate co-precipitates with the calcium sulfite, resulting in a mixed crystal (or solid solution) of calcium-sulfur salts. Gypsum is not formed. The relatively high sulfite concentrations in the solution prevent soluble calcium concentrations from reaching the levels required to exceed the gypsum solubility product, and the system operates unsaturated with respect to calcium sulfate (as gypsum).

The amount of sulfate co-precipitated with the calcium sulfite is a function of the concentrations of sulfate and sulfite and the reactor solution pH. As the concentration of sulfate increases relative to sulfite, the amount of sulfate precipitation increases. Thus, as the rate of oxidation increases, the ratio of sulfate to sulfite in solution will increase until the rate of calcium sulfate precipitation is sufficient to keep up with the rate of sulfate formation by oxidation. Co-precipitation enables the system to keep up with oxidation rates equivalent to 25% to 30% of the SO_2 absorbed without intentionally purging sodium sulfate. Under such conditions, the solution will remain unsaturated with respect to calcium sulfate, thereby avoiding high soluble calcium concentrations and attendant scaling problems. The sulfate/sulfite ratio can obviously be increased by increasing sulfate concentrations in the liquor and/or by decreasing the sulfite or active sodium concentration. However, it is desirable to operate with active sodium concentrations above about 0.15 M, below this concentration the system becomes "dilute" in active sodium--a condition in which the system can become saturated or supersaturated in calcium sulfate with soluble calcium concentrations rising to levels of 400-800 ppm. The result would be higher scale potential and a general deterioration in cake properties.

In practice, a concentrated-mode system would normally be operated at a total sodium concentration consistent with the highest sustained level of oxidation expected--that is, a sodium concentration high enough that under the worst conditions, with regard to oxidation, the active sodium concentration would not fall below about 0.15 M Na^+. The liquor composition can then vary with changes in oxidation, the sulfate/sulfite ratio adjusting to whatever level is required to keep up with sulfate formation in the system.

3. Solids Dewatering

Solids dewatering is a purely mechanical process involving thickening of the reactor effluent slurry to 15-30 wt % solids, followed by filtration to produce a waste filter cake. While the cake is being formed, it is washed with fresh water to recover sodium salts that would otherwise be lost in the process liquor discharged with the cake. The filter cake represents the only waste discharged from the process. There are no other purges from the system.

The solids content and chemical composition of the waste cake produced from SO_2 control alone will depend primarily upon the amount of oxidation, the quality of raw lime, and the chloride content of the coal. In general, for utility boilers firing medium to high sulfur coal, the waste cake would normally be expected to contain 55-70 wt % insoluble solids and have a chemical composition typically within the following ranges (dry basis):

$CaSO_3 \cdot 1/2\ H_2O$ = 80-85 wt %

$CaSO_4 \cdot 1/2\ H_2O$ = 5-15 wt %

$CaCO_3$ + Inerts = 5-10 wt %

Soluble Sodium Salts = 1-3 wt %

The amount of sodium lost in the cake will depend primarily upon the total sodium concentration in the process liquor (which is largely a function of the amount of oxidation and the chloride content of the coal), and the extent of cake washing. The minimum level of sodium losses for a concentrated mode system is generally on the order of 0.5-1.0 wt.% of the total dry solids, approximately 0.01 to 0.02 moles of Na^+/mole SO_2 scrubbed. This minimum level is dictated by liquor which is occluded within calcium-sulfur crystals or trapped in interstices of agglomerates which cannot be practically washed from the waste; and the rate at which chlorides and other highly soluble anionic species enter the system. Such highly soluble species must be purged in the cake, as in any nonrecovery solid waste producing process, at the rate at which they enter the system. In a sodium-based dual alkali system, these would leave as soluble sodium salts (e.g., NaCl) rather than as soluble calcium and magnesium salts.

4. Raw Materials Preparation

Only two chemicals are required for the operation of the dual alkali system: lime for regeneration and soda ash to make up for sodium losses.

Typically, lime would be slaked and hydrated prior to being fed to the reactor system. At the Cane Run Station, however, carbide lime is available to LG&E as a slurry containing 25-30% insoluble solids. The slurry is pumped to a day tank from which lime slurry is fed to the primary reactors.

The rate of addition of soda ash to the system is adjusted to compensate for the losses of sodium in the cake. Typically, the soda ash makeup requirement should amount to about 1-3% of the total alkali requirement on a molar basis (includes the alkali required for both SO_2 control and chloride absorption/neutralization).

Clarified liquor from the thickener hold tank is added to soda ash to prepare the makeup solution in the soda ash tank. The resulting solution may be fed to either the thickener or the absorbers. If added to the thickener center well, the sodium carbonate will soften the regenerated liquor, reacting with dissolved calcium and precipitating calcium carbonate. The precipitation of calcium carbonate will cause a slight loss in the overall utilization of calcium since it is not removing any sulfur from the system. The calcium loss, however, is small and amounts to about 1% or less of the calcium fed to the system.

Alternatively, the soda ash makeup solution can be added directly to the absorber. The sodium carbonate reacts with acidic sodium bisulfite in the absorber, producing sodium sulfite and liberating CO_2. In this manner, the soda ash is used directly in the absorption of SO_2 and thus avoids any small, unnecessary loss of calcium due to calcium carbonate precipitation.

At LG&E, the soda ash solution will be directed to the absorber. The flexibility exists, however, for pumping the makeup solution to the thickener center well.

B. POLLUTION CONTROL CAPABILITIES

1. SO_2 Control

The sodium-based dual alkali process, operating in the concentrated active sodium mode, is capable of SO_2 removal efficiencies in excess of 95% over any range of inlet SO_2 concentrations encountered in coal-fired utility boiler applications. In most cases, removal efficiencies approaching 99% can be achieved on a continuous basis, as was demonstrated during the test program on a 20 Mw dual alkali prototype system[1]. These high efficiencies are accomplished by proper selection and design of the absorber unit

[1] LaMantia, C.R., et. al., "Final Report: Dual Alkali Test and Evaluation Program", Volume III, EPA Contract 600/7-77-050, May 1977.

and by adjustment of the active sodium/SO_2 stoichiometry in the absorber. Such variation in SO_2 absorption efficiency can be effected without influencing the overall lime stoichiometry or the sodium makeup requirement. These high SO_2 removal efficiencies can be achieved in tray-type absorbers at low L/G ratios, typically in the range of 5-10 gpm/1,000 acfm of saturated gas. The pressure drop across the trays may be minimal, in the range of 4-6 inches WG.

The high SO_2 removal capability of this process, when used in conjunction with a boiler equipped with adequate control of particulate matter, allows the option of removing virtually all of the SO_2 from the flue gas treated in the scrubber and bypassing hot, untreated gas to provide part or all of the reheat while still meeting the overall plant SO_2 emission regulations in the combined treated and untreated flue gas. In such a system, the scrubber size can be reduced since not all flue gas is treated and the reheat requirements are reduced or eliminated.

2. Control of Particulate Matter

Removal of particulate matter can be accommodated in the process by appropriate selection of scrubbers. If particle removal is to be accomplished as part of the overall system, then a higher energy particulate matter removal device, such as a venturi scrubber, may be incorporated in this system to provide for removal of both SO_2 and particulate matter. Removal of particulate matter down to 0.02 grains/scfd or lower can be accomplished using venturi scrubbers at moderate pressure drops on the order of about 20 inches WG.

3. Chloride Control

A major fraction of the chlorides in coal (greater than 90%) is volatilized and appears in the flue gas as HCl. Any aqueous-based scrubbing system would be highly effective in absorption of HCl (and any HF) in the flue gas. As a result, chloride concentrations will build in the closed liquor loop to levels such that the rate at which chloride is discharged from the system in the washed cake will equal the rate at which chloride enters the system with the flue gas. Steady-state levels of chloride in the closed liquor loop of a 20 Mw prototype CEA/ADL dual alkali system rose to as high as 11,000 ppm (0.05-0.1% Cl in coal) with no apparent effect on the process operation[1]. Tests of the lime regeneration reaction at ADL have shown that lime utilization and solids properties are unaffected by chloride concentration as high as 25,000 ppm.

[1] Ibid.

IV. DESCRIPTION OF THE DUAL ALKALI PROCESS APPLICATION AT THE CANE RUN STATION

This chapter provides a description of the application of the dual alkali technology to the specific flue gas desulfurization requirements for LG&E's Cane Run Unit No. 6. In designing the system, consideration was given to the sources, characteristics, and amounts of flue gas to be cleaned as well as to the type and sources of the required raw materials.

A. BOILER SYSTEM DESCRIPTION

Cane Run Unit No. 6 consists of a pulverized coal-fired steam boiler, built by Combustion Engineering, with a Westinghouse turbine-generator. The unit operates from a minimum of 60 Mw during off-peak hours to a maximum load of 300 Mw during peak hours. The annual average load is equivalent to approximately 180 Mw (about 60% of the gross peak capacity).

Flue gas from the boiler passes, in parallel streams, through two Ljungstrom combustion air preheaters. Each air preheater discharges flue gas through separate ducts to an electrostatic precipitator designed for 99.4% removal efficiency of particulate matter (weight basis). From the precipitator, the gasses enter two parallel induced draft fans, each handling 50% of the total gas.

The sulfur dioxide removal system, installed between the existing induced draft fans and the stack, draws hot flue gas from the outlet of the induced draft fans through two booster fans. The scrubbed gas is reheated and then returned to the existing entrance to the stack. Appropriate dampers have been provided to allow bypass of gas around the scrubber system using the existing ductwork.

Coal for Unit No. 6 is received from a number of sources. A dry ultimate analysis typical of the coal fired is given in Table IV-1. The average sulfur content on a dry basis is 4.8% and varies from 3.5% to 6.3%. The average chloride content of the coal is 0.04% and varies from 0.03% to 0.06%. The average 4.8% sulfur content and 11,000 Btu/lb will result in an SO_2 emission level equivalent to about seven times that allowed by the present Federal New Source Performance Standards (1.2 lbs of SO_2/MM Btu).

B. DESIGN CONDITIONS FOR THE DUAL ALKALI SYSTEM

The design basis for the dual alkali system is summarized in Table IV-2. Design conditions correspond to coal containing 5% sulfur and 0.04% chloride and having a heating value of 11,000 Btu/lb on a dry basis.

TABLE IV-1

ULTIMATE ANALYSES OF COAL FIRED IN UNIT NO. 6

(Dry Basis)

	Typical Analysis, %	Range, %
Carbon	67.15	64.0 - 70.0
Hydrogen	4.72	4.3 - 5.25
Nitrogen	1.28	0.6 - 1.5
Chloride	0.04	0.03 - 0.06
Sulfur	4.81	3.5 - 6.33
Ash	17.06	15.5 - 24.5
Oxygen	4.94	3.8 - 6.2
	100.00	
Moisture	8.95	8.0 - 10.75
Heat Content, Btu/lb dry coal	11,000	9,500-12,400 Maximum 10,400-11,900 Normal

TABLE IV-2

DESIGN BASIS

Coal (Dry Basis):	
Sulfur	5.0% S
Chloride	0.04% Cl^-
Heat Content	11,000 Btu/lb
Inlet Gas:	
Flow Rate (Volumetric)	1,065,000 acfm
(Weight)	3,372,000 lb/hr
Temperature	300°F
SO_2	3,471 ppm (dry basis)
O_2	5.7 vol.%
Particulate Matter	$\leq$0.10 lb/10^6 Btu
Outlet Gas:	
SO_2	$\leq$200 ppm ($\sim$0.45 lb/10^6 Btu)
Particulate Matter	$\leq$0.10 lb/10^6 Btu

The flue gas conditions at the inlet of the dual alkali system are given in Table IV-3. The design gas capacity of 3,372,000 lbs of flue gas per hour (combined flow to both scrubbers) corresponds to the boiler peak load capacity of 300 Mw.

The dual alkali system is designed to meet all applicable federal, state, and local pollution control and safety regulations. The maximum SO_2 concentration in the scrubbed gas will be 200 ppm (for coal containing up to 5% sulfur), well below requirements of the current NSPS. There will be no discharge of process liquor from the system; and the disposal of the waste solids produced will meet all applicable federal, state, and local solid waste disposal regulations currently in effect. None of the wastes will be discharged to or allowed to enter any naturally occurring surface water. Plans for disposal are discussed as a part of this report.

The existing 518 foot stack will be the only source of gaseous emissions from the system. A flue gas bypass will be provided to allow untreated boiler flue gas to enter the stack, bypassing the dual alkali system. The scrubber system is designed to be isolated from the flue gas during periods in which the bypass is open to allow safe entry into the scrubber system for maintenance and inspection while the boiler continues normal operation. Also, each absorber can be isolated independently and maintenance can be provided to one absorber while the other absorber is in operation. The duct dampers are designed such that with the dual alkali system in operation and the bypass closed, no more than 1.0% of the total flue gas will leak through the bypass system into the stack.

The dual alkali system has been equipped with sufficient instruments, in addition to those required to operate the process, to permit accurate measurements of the appropriate streams required to calculate material and energy balances. In particular, instrumentation has been provided to permit continuous monitoring of SO_2 concentrations in the flue gas entering and leaving the control system as well as the measurement of the quantities of chemicals and water entering the system and filter cake discharge.

C. GUARANTEES

The process guarantees listed below include minor revisions since publication of the project manual containing the preliminary design and cost estimate (EPA-600/7-78-010, January, 1978).

1. Sulfur Dioxide Emission

The system shall provide such control that emissions from the stack shall be no greater than 200 ppm by volume SO_2 dry basis not including SO_2 added from the operation of the reheaters and without additional air dilution when burning the coal containing less than 5% sulfur. When burning coal containing 5% sulfur or greater, the system shall remove at least 95% of the sulfur dioxide in the inlet flue gas.

TABLE IV-3

FLUE GAS CONDITIONS AT THE INLET OF THE DUAL ALKALI SYSTEM

Normal Operating Temperature	300°F
Maximum Gas Temperature for Periods up to 5 Mins.	600°F
Normal Pressure at I.D. Fan Outlet	-1" to +2" WG
Boiler Excess Air	25% Maximum 35%
Air Heater Leakage	10%
Flue Gas Density at Sea Level @ 70°F	0.078 lb/ft^3
Total Pressure at Stack Entrance	+2" WG
Boiler Load Points (lbs/hr flue gas):	
Design	3,372,000
Boiler Maximum Continuous Rating	3,003,000
Control Load	1,440,000
Minimum Normal Operating Load	658,000

2. Particulate Matter Emission

In addition to meeting applicable regulations, the system shall also meet Federal New Source Performance Standards for emissions of particulate matter under all conditions of boiler operation. The dual alkali system shall not add any particulate matter to the emissions of particulate matter that is received by the system from the LG&E Cane Run Unit No. 6 electrostatic precipitator.

3. Lime Consumption

The consumption of lime in the system shall not exceed 1.05 moles of available CaO in the lime feed per mole of SO_2 removed from the flue gas.

4. Sodium Carbonate Makeup

Soda ash makeup shall not exceed 0.045 moles of Na_2CO_3 per mole of SO_2 removed from the flue gas provided that the chloride content of the coal burned averages 0.06% or less. If the average chloride content of the coal is above 0.06%, then additional sodium carbonate consumption will be allowed at the rate of 1/2 mole Na_2CO_3 for each mole of chloride (Cl^-) in the flue gas resulting from chloride in excess of 0.06% in the coal. The Seller as part of the guarantees shall perform the necessary research and design to reduce the makeup requirements of Na_2CO_3 from the guarantee point to a level approaching minimal makeup.

5. Power Consumption

At the peak operating rate (300 Mw), the system shall consume a maximum of 1.2% of the power generated by the unit.

6. Waste Solids Properties

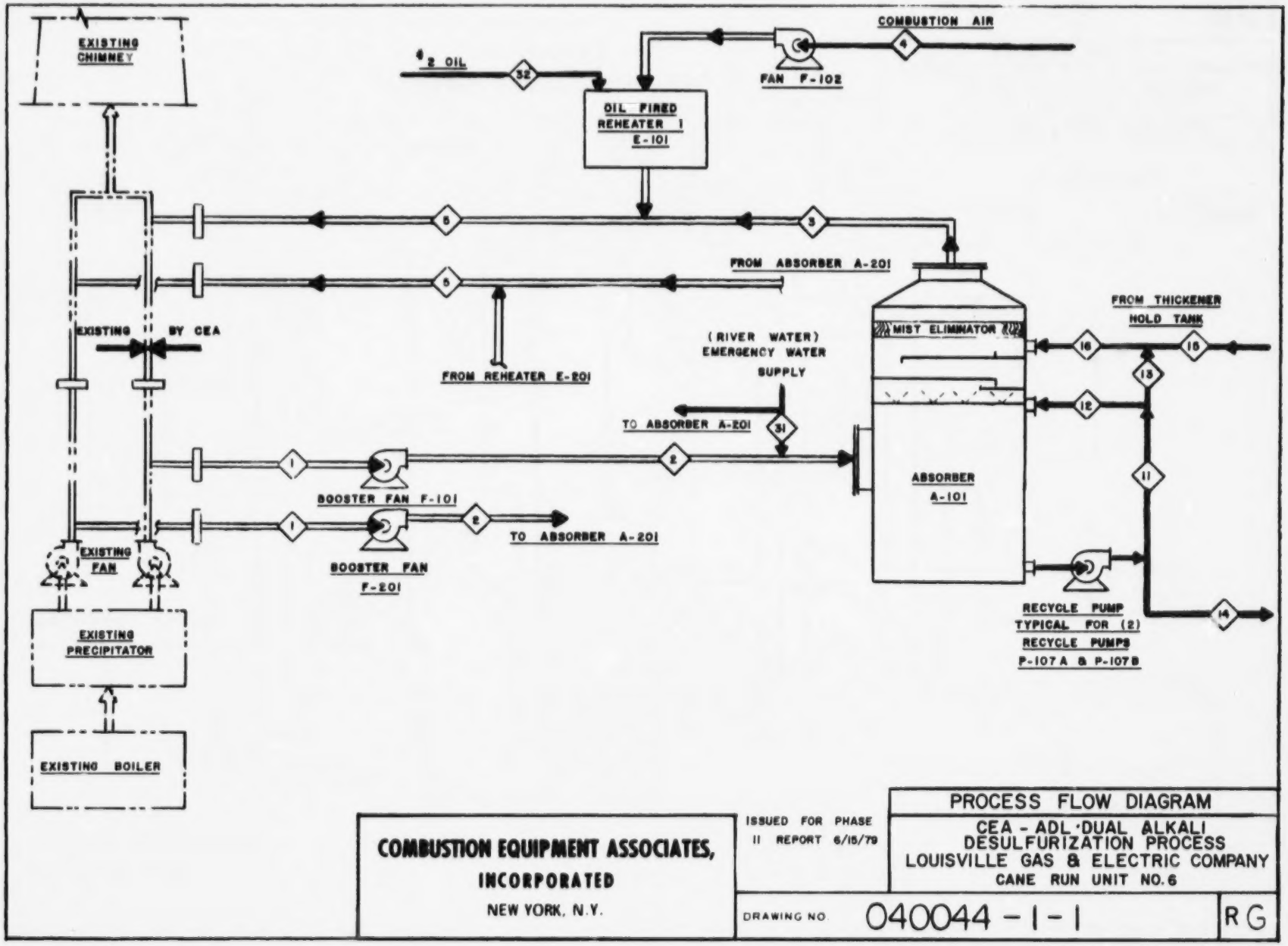
The waste produced by the vacuum filter shall contain a minimum of 55% insoluble solids.

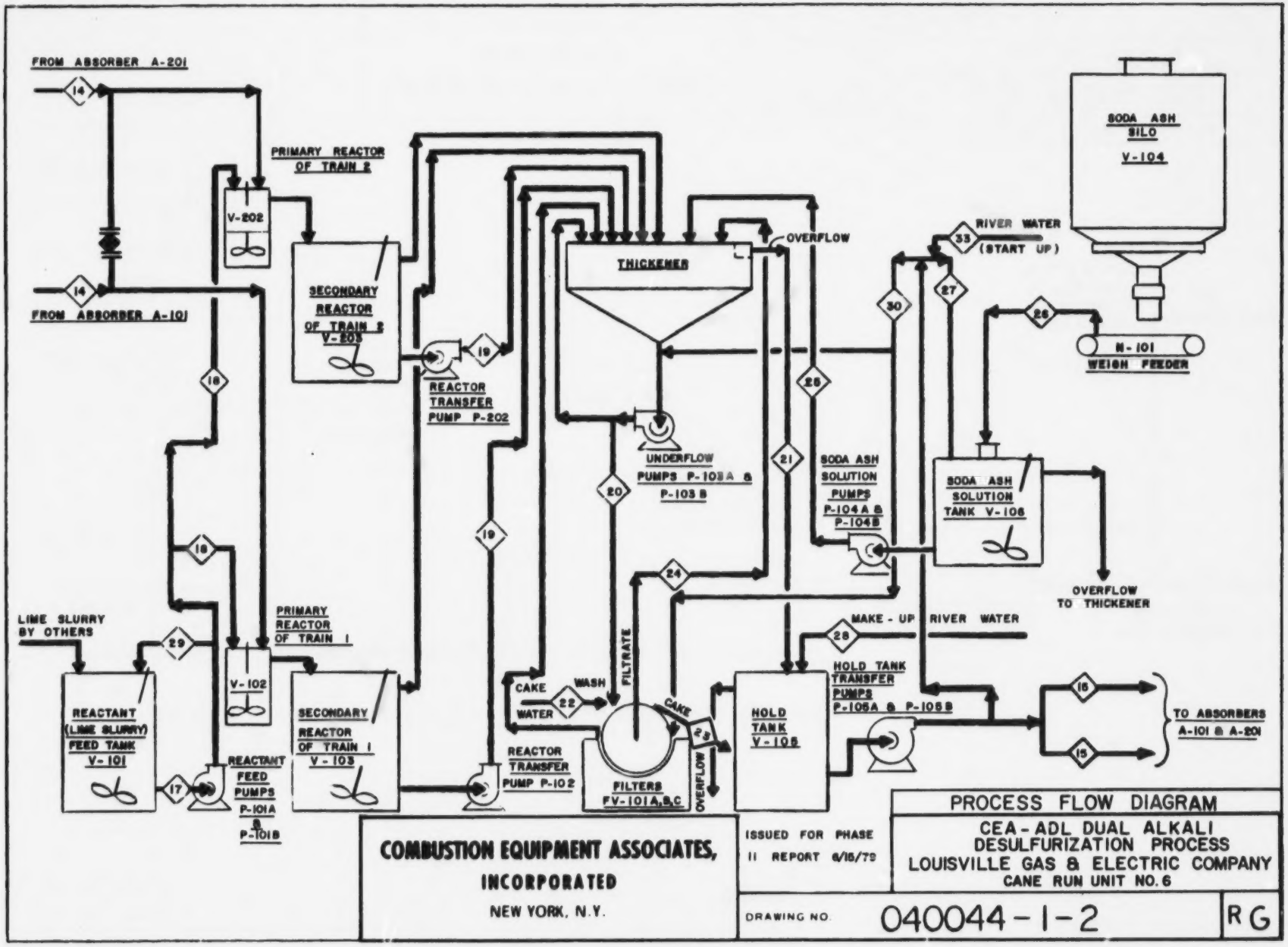
7. SO_2 System Availability

The system shall have an availability (as defined by the Edison Electric Institute for power plant equipment) of at least 90% for one year. Thus, the system shall be available for operation at least 90% of the calendar time.

D. PROCESS DESCRIPTION

The description of the dual alkali system for Cane Run Unit No. 6 can be conveniently divided into six parts: (1) absorber section (flue gas scrubbing); (2) reactor section (regeneration of scrubbing solution); (3) solids dewatering; (4) raw materials preparation; (5) waste disposal; and (6) provisions for spills and leaks. The process flow diagram for the overall system is shown in CEA Drawings 040044-1-1 and 040044-1-2.


EXISTING CHIMNEY
#2 OIL
32
COMBUSTION AIR
4
FAN F-102
OIL FIRED REHEATER 1 E-101
5
3
FROM ABSORBER A-201
5
FROM THICKENER HOLD TANK
EXISTING
BY CEA
FROM REHEATER E-201
(RIVER WATER) EMERGENCY WATER SUPPLY
MIST ELIMINATOR
16
15
18
12
TO ABSORBER A-201
31
1
2
BOOSTER FAN F-101
ABSORBER A-101
11
1
2
TO ABSORBER A-201
EXISTING FAN
BOOSTER FAN F-201
RECYCLE PUMP TYPICAL FOR (2) RECYCLE PUMPS P-107A & P-107B
14
EXISTING PRECIPITATOR
EXISTING BOILER
COMBUSTION EQUIPMENT ASSOCIATES, INCORPORATED
NEW YORK, N.Y.
ISSUED FOR PHASE II REPORT 6/15/79
PROCESS FLOW DIAGRAM
CEA - ADL DUAL ALKALI DESULFURIZATION PROCESS
LOUISVILLE GAS & ELECTRIC COMPANY
CANE RUN UNIT NO. 6
DRAWING NO. 040044 - 1 - 1
R G


FROM ABSORBER A-201
14
PRIMARY REACTOR OF TRAIN 2
V-202
SECONDARY REACTOR OF TRAIN 2 V-203
14
FROM ABSORBER A-101
18
19
REACTOR TRANSFER PUMP P-202
THICKENER
OVERFLOW
33
RIVER WATER (START UP)
SODA ASH SILO V-104
30
27
26
N-101 WEIGH FEEDER
25
UNDERFLOW PUMPS P-103A & P-103B
21
SODA ASH SOLUTION PUMPS P-104A & P-104B
SODA ASH SOLUTION TANK V-106
20
19
24
OVERFLOW TO THICKENER
MAKE - UP RIVER WATER
28
HOLD TANK TRANSFER PUMPS P-105A & P-105B
HOLD TANK V-105
16
15
TO ABSORBERS A-101 & A-201
LIME SLURRY BY OTHERS
29
PRIMARY REACTOR OF TRAIN 1
V-102
REACTANT (LIME SLURRY) FEED TANK V-101
SECONDARY REACTOR OF TRAIN 1 V-103
17
REACTANT FEED PUMPS P-101A & P-101B
CAKE WASH WATER
22
FILTRATE
CAKE
23
OVERFLOW
REACTOR TRANSFER PUMP P-102
FILTERS FV-101A,B,C
COMBUSTION EQUIPMENT ASSOCIATES, INCORPORATED
NEW YORK, N.Y.
ISSUED FOR PHASE II REPORT 6/15/72
PROCESS FLOW DIAGRAM
CEA - ADL DUAL ALKALI DESULFURIZATION PROCESS
LOUISVILLE GAS & ELECTRIC COMPANY
CANE RUN UNIT NO. 6
DRAWING NO. 040044-1-2
RG

1. Absorber Section

The absorber section consists of two identical scrubber modules. Each module is made up of a booster fan, an absorber, an oil-fired flue gas reheater, and two recirculation pumps (one operating and one back-up).

Flue gas is drawn from the existing induced draft fans and is forced through the absorbers by means of the booster fans. The basic design and control philosophy is predicated on both absorbers operating simultaneously with each handling half the boiler load. However, during periods of low boiler load, one absorber is capable of handling all of the gas. Hence, in order to allow for greater flexibility in operation and provide for the possibility of allowing maintenance on one module while the system is still in service, a common duct connecting the two booster fan inlets has been incorporated. A bypass is also provided to allow complete shutdown of the scrubbers while the boiler is still on line. The bypass and FGD system inlet dampers have been properly interlocked to enable bringing the absorbers on- or off-line without interruption of the boiler operation.

Hot flue gas (~300°F) entering each absorber is first cooled by a set of sprays which direct scrubbing solution at the underside of the bottom tray. In addition to providing temperature control at the bottom of the absorbers, these sprays keep the underside of the tray and the bottom of the absorber free of any buildup of fly ash solids. The cooled, saturated gas passes through a set of two trays where SO_2 is removed and then through a chevron-type demister. After leaving the absorber, the scrubbed gas is reheated 50F° (to a temperature of about 175°F) by mixing it with hot flue gas from an oil-fired reheater to avoid condensation and corrosion as it is exhausted to the stack.

The scrubbing solution which flows counter-current to the gas is collected at the bottom of the absorber. This liquor is used for the quench sprays and as a recycle stream to the top tray for pH control. Since the scrubbing solution is regenerated with lime, the feed to the absorber contains sodium hydroxide and therefore is very alkaline. A high tray feed pH increases the absorption of CO_2 which in turn increases the potential for $CaCO_3$ scaling in the absorber. The recycle also ensures proper liquid loading on the trays. A bleed stream from the bottom of the absorber is sent to the reactor system for regeneration. Provisions have been made for an automatic emergency water supply to spray and quench the hot flue gas in the event that the recycle pump fails.

Pressures and temperatures are measured at appropriate points throughout the gas and scrubber systems. Removal of SO_2 is monitored by continuous SO_2 analyzers at the inlet and outlet of the absorber. The pH of the bleedstream and of the feed to the top tray is continuously monitored also.

2. Reactor Section

This section of the system consists of two identical reactor trains each containing a primary and secondary reactor.

At operation under design conditions, each reactor train would handle the regeneration of solution from the corresponding absorber; although for short periods of time, a single reactor train is capable of handling the solutions from both absorbers operating at design conditions. When the boiler is firing typical or average coal (3.5-4.0% sulfur), only one reactor train is normally required.

The spent scrubbing solution is introduced to the primary reactor along with slurried carbide lime from the lime day tank. In this short residence time reactor (3-15 minutes), the regeneration of scrubbing solution begins. The primary reactor overflows into a second, longer residence time reactor (30-60 minutes) where regeneration is completed. The secondary reactors are maintained at a pH typically in the range of 11 to 12 by controlling the amount of lime slurry fed to the primary reactors.

The slurry from the secondary reactor is fed to the thickener feed well where the separation of the regenerated solution from the solid waste is initiated.

3. Solids Dewatering

The reactor effluent, a slurry containing 2-5% insoluble solids, is directed to the feed well of the thickener. The thickener is generally operated to provide an underflow (thickened) slurry containing 15-30 wt.% solids, even though the slurry can be thickened to 40 wt.% solids or more. Slurries in the range of about 25 wt.% solids allow better control of filter cake washing, control which cannot always be achieved with variations in filter drum speed and pool depth. The thickener underflow slurry is recirculated past the filters in a recycle loop that returns the slurry to the solids zone of the settler. A bleed from this recirculation loop is fed to the filters. Each filter is equipped with an overflow pipe returning to the solids zone in the thickener to allow for operation in an overflow mode and thereby provide against inadvertent overflow of the filter hold tank.

There are three filters, each rated to handle 50% of the total solids produced at the design conditions. Each filter can be operated independently. For optimum performance (to obtain cake containing high dry solids and low soluble salts) it is desirable to operate the filters at fixed conditions (constant drum speed, submergence, wash ratio, etc.). Therefore, the cake rate is controlled by changing the number of filters in operation. The number of filters in operation is determined by the

quantity of solids accumulated in the thickener, which is reflected in the solids concentration in the underflow slurry and the thickener rake lift position indicator. The density of the underflow slurry is measured and thickener hold tank liquor is added as required to maintain the percent solids in the underflow slurry at about 20-25%. The number of filters in operation is changed if the concentration of solids in the underflow slurry cannot be controlled using the dilution liquor. For operation at typical conditions (3.5% S and an average daily load of 60-70%), it is anticipated that only 2 filters need to be operated about one shift per day.

The solid cake is washed on the filter using a series of water spray banks. This wash removes a large fraction (approximately 90%) of the occluded soluble salts from the cake and returns these salts to the system, thereby reducing sodium losses and minimizing sodium carbonate makeup. The total wash rate will usually be set to be a constant percentage of the cake discharge rate. The mixed filtrate and wash liquor from the filter is collected in the filtrate sump from which it is returned to the thickener.

Clear liquor overflow from the thickener is collected in the thickener hold tank which both provides surge capacity for the absorbent liquor feed to the scrubber system and maintains overall control of the volume of liquor in the system. Water is added to this hold tank to make up for the difference between total system water losses (evaporation and cake moisture) and total water inputs from other sources (sodium makeup solution, pump seals, lime feed, and cake wash).

4. Raw Materials Preparation

Two chemicals are required for the operation of the dual alkali system: lime for absorbent regeneration and soda ash to make up for sodium losses.

Carbide lime, a byproduct of acetylene production, is available to LG&E at a significantly lower price than commercial lime. The carbide lime is barged to the Cane Run Station as a slurry containing approximately 30% insoluble solids. The slurry is hydraulically unloaded and pumped to a main storage tank at the plant. Excess liquor is then decanted from the tank and returned for reuse in transport of carbide lime. From the main storage tank, the lime slurry will be pumped to a grinding system consisting of a hydroclone and wet ball mill to prevent feeding of oversized material. From the grinding system, the lime will be pumped to the dual alkali system day tank which supplies lime to the primary reactor at the appropriate rate. Since the installation of the grinding equipment has been delayed beyond the startup date for the dual alkali system, a disintegrator with coarse screens has been temporarily installed upstream of the day tank for rough sizing of the raw carbide lime until the permanent grinding facility is completed.

The utilization of available $Ca(OH)_2$ in the slurry (conversion to $CaSO_x$ salts) is expected to be on the order of 98%. Thus the overall feed stoichiometry to the system (moles of available $Ca(OH)_2$/mole of SO_2 absorbed) should be on the order 1.0 (taking into account the soda ash makeup). The specifications for the carbide lime as delivered to the dual alkali system are given in Table IV-4. The carbide lime contains 92.5 wt.% $Ca(OH)_2$ and has a particle size distribution equivalent to 90% through a 325 mesh screen.

The addition of sodium carbonate to the system is to compensate for the losses of sodium in the cake. Despite washing the cake, some liquor, containing soluble sodium salts, will inevitably remain occluded in the cake.

Dry, dense soda ash is received at the plant and stored in the soda ash silo from which it is fed to the soda ash solution tank by means of a weigh feeder. The rate of addition is proportional to the amount of lime slurry fed to the system. Such relationship arises from the fact that the sodium losses in the cake can be expressed as a percentage of the solids produced, which in turn are related to the amount of lime added to the system.

Soda ash solution can be made up using either clarified liquor drawn from the thickener hold tank or fresh water. Provisions have also been made to feed the soda ash solution either directly to the absorbers or to the thickener. The normal mode of operation is to prepare the soda ash makeup solution using clarified liquor and to feed it to the absorbers.

5. Waste Disposal

A long-range plan for the disposal of the dual alkali filter cake has been developed as a part of an overall disposal plan for all FGD wastes produced at the Cane Run Station. The plan involves stabilization of the wastes generated by each of the three FGD systems via the addition of lime and fly ash and dry landfill of the stabilized material adjacent to the plant. Waste from the dual alkali system on Unit No. 6 will be handled, processed, and disposed of independently of the wastes produced by the direct lime scrubbing systems on Unit Nos. 4 and 5. Wastes from these latter two units will be combined and handled in a common processing plant.

The waste processing plant will be installed at the end of the filter building. Filter cake from the dual alkali system is discharged into a single conveyor (the filters are arranged in series--end to end), which will carry the cake directly to the pug mill mixer in the processing plant. A belt weigh element near the end of the conveyor measures the quantity of cake produced. Fly ash from Unit No. 6 will be pneumatically transported to the processing plant and stored in a live-bottom silo, from which it will be discharged to a screw conveyor and fed to the pug mill. The ash feed will be adjusted according to the filter cake rate.

TABLE IV-4

CARBIDE LIME SPECIFICATIONS

	Carbide Lime Slurry[a]
Calcium hydroxide $Ca(OH)_2$	92.50
Available calcium oxide CaO	70.01
Calcium carbonate $CaCO_3$	1.85
Silica SiO_2	1.50
Iron and alumina oxides R_2O_3	1.60
Magnesium oxides MgO	0.07
Sulfur	0.15
Phosphorus	0.01
Free carbon	0.25
Free Water	--
Not analyzed	2.07

[a]Available as slurry containing 30% insoluble solids.

Source: Airco catalog (1969).

Carbide lime will also be fed (as a 30% slurry) to the pug mill at a rate equivalent to a few percent of the combined dry weight of ash and cake. The carbide lime will be stored in a small day tank in the processing plant which will be charged from the main plant storage tank.

The pug mills will discharge to radial stacking conveyors. The stackers will spread the material in a windrow (interim stockpiling area) where it will remain for one to three days to initiate the stabilization reactions. The material will be removed from the windrow and loaded into rear dump trucks using a front-end loader. The trucks will then transport the material to a dry landfill south of the plant on LG&E property. Provisions have also been made to allow direct loading of the trucks from the stacking conveyors should this mode of operation prove satisfactory.

Installation of the waste processing plant is scheduled for completion in the fall of 1979. In the interim, the filter cake conveyor will discharge through a feed chute directly into trucks which will transport the waste to a temporary storage area. Once the processing plant is operational, this waste will be reclaimed from the storage area and processed along with fresh filter cake.

6. Provisions for Spills and Leaks

Filter cake is the only product of the dual alkali system. The system will be operated in a closed loop and there will be no other solid or liquid discharge from the system. In order to avoid inadvertent discharge of any process liquor, a number of provisions have been made in the process design.

(a) Drains are provided at all pump stations and inside the filter building to direct all process pump seal water leaks, pump and piping flush water and equipment and building washdown water to the filtrate sump for return to the thickener along with the filtrate and cake wash water.

(b) A separate sump is provided to collect vacuum pump seal water. This is noncontact water. However, should the vacuum pump seal water become contaminated with process liquor, provisions have been made to also pump this water to the thickener.

(c) The emergency overflow from the absorbers are also drained to the filtrate sump for return to the system.

(d) The thickener has approximately three feet of sidewall above the overflow weir instead of the usual 6-8 inches; and the thickener hold tank has a height equal to that of the thickener. The additional sidewall height in the thickener and hold tank is to allow for temporary storage of liquor from the other process vessels in the system if required during maintenance;

and to prevent any short-term spills of liquor due to water imbalances resulting from extreme process upsets. The total capacity that this additional height provides is equivalent to the total capacity of all other process vessels in the system.

(e) The thickener hold tank is equipped with an emergency overflow tank which will begin to fill when extreme levels have been reached in the thickener and hold tank. Filling of this tank will indicate the need for immediate corrective action.

E. MATERIAL BALANCES

Cane Run Unit No. 6 has a rated capacity of 280 Mw and a peak load capacity of 300 Mw. The system has been designed to handle the peak load, and the material balances presented here have been calculated for the design conditions unless otherwise specified. The basis for the material balances is given in Table IV-5. The design coal contains 5% sulfur and 0.04% chloride on a dry basis. All estimates are based on 94% of the sulfur in the coal appearing in the flue gas.

1. Overall Material Balance

The overall material balance for the dual alkali system operating at design conditions is given in Table IV-6. Removal of 94.2% of the 390 lbs/min of SO_2 present in the inlet gas to the scrubbing system generates 1,246 lbs/min of waste cake. The moisture content of this cake is about 36%. Lime (as $Ca(OH)_2$) is added to the system at a rate of 460 lbs/min on a dry basis. At 30% insoluble solids in the lime slurry, the slurry feed rate would equal 1,535 lbs/min or a total of about 150 gpm. Soda ash is added at a rate of 13.7 lbs/min and process water at a rate of 369 gpm.

An overall water balance at design conditions is given in Table IV-7. Water enters the system in the lime slurry (both as slurry water and chemically combined water), cake wash, instrument purge, and pump seals. All the water used for pump seals eventually enters the system. The water that drains out of the seals is collected in sumps and returned to the thickener.

More than 75% of the water lost from the system at design conditions is due to evaporation of water to the flue gas. The amount of water evaporated in the absorbers depends on the boiler load and the temperature and humidity of the inlet flue gas. Water is also removed from the system with the waste cake, both as free water and as chemically bound water. In addition to these losses, which are directly related to process operations, water enters and leaves the system via liquid surfaces exposed to the atmosphere (principally the thickener and thickener hold tank). Based on meteorological conditions at Louisville, thickener and hold tank configurations and average liquor temperatures, it is anticipated that

TABLE IV-5

BASIS FOR MATERIAL BALANCES AT DESIGN CONDITIONS

Coal (dry basis):

Sulfur	5.0% S
Chloride	0.04% Cl
Heating value	11,000 Btu/lb
Sulfur volatilized	94% of S in Coal

Inlet Gas:

Flow rate (volumetric)	1,065,000 acfm
(weight)	3,372,000 lb/hr
Temperature	300°F
SO_2	3,471 ppm (dry basis)
O_2	5.7 vol. %
Particulate Matter	$\leq$0.1 lb/10^6 Btu

Outlet Gas:

SO_2	200 ppm (dry basis) ($\sim$0.45 lb/10^6 Btu)
Particulate Matter	$\leq$0.10 lb/10^6 Btu

Absorber Feed Concentration:

Na^+ associated with	
OH^- and $CO_3^=$	0.09 M
$SO_3^=$	0.36 M

Oxidation and Sodium Makeup Rates:

Oxidation	10% ΔSO_2 (molar basis)
Na_2CO_3	0.045 moles Na^+/mole ΔSO_2 (criteria: $\leq$0.0495)

Calcium Feed:

Solids in slurry	30%
Available $Ca(OH)_2$	92.5% (weight basis)
$Ca(OH)_2$ utilization	98% (equivalent to 1.003 moles available $Ca(OH)_2$/mole ΔSO_2; criteria: $\leq$1.01)

Waste Solids:

Wash ratio	2.0 displacement washes
Insoluble solids	63 wt.% (criteria: $\geq$55 wt.%)

TABLE IV-6

OVERALL MATERIAL BALANCE AT DESIGN CONDITIONS

Basis: Coal - 5.0% sulfur
- 0.04% chloride
- 11,000 Btu/lb

Full load (300 megawatts)
FGD inlet SO_2 - 390 lbs/min
SO_2 removal - 94.2%

	Consumption Rate	Lbs/lb Coal Fired	Lbs/lb SO_2 Absorbed
Makeup Materials			
Process Water	369 gpm	---	---
Lime (92.5 wt.% $Ca(OH)_2$, dry basis)	460 lbs/min	0.111	1.25
Soda Ash	13.7 lbs/min	0.003	0.037
Cake Production			
Dry Basis	804 lbs/min	0.194	2.18
Wet Basis	1,246 lbs/min	0.300	3.39

TABLE IV-7

OVERALL WATER BALANCE AT DESIGN CONDITIONS

	Inlet (gpm)	Outlet (gpm)
Process Water		
• Lime Slurry - Free Water	128.9	
- Chemically Combined	12.4	
• Filter Cake Wash	90.7	
• Pump Seal	20.0	
• Instrument Purge	0.5	
• Rain	0.6	
• Make-up (to Thickener Hold Tank)	116.2	
• Evaporation - Absorber		278.1
- Open Tanks		32.0
• Filter Cake - Free Water		53.1
- Chemically Combined		6.1
	369.3	369.3
Non-Contact Water		
• Fan Bearing Cooling	175	
• Vacuum Pump Seal	40	
• Fan Bearing Cooling		175
• Vacuum Pump Seal		40
	215	215

TABLE IV-6

OVERALL MATERIAL BALANCE AT DESIGN CONDITIONS

Basis: Coal - 5.0% sulfur
- 0.04% chloride
- 11,000 Btu/lb

Full load (300 megawatts)
FGD inlet SO_2 - 390 lbs/min
SO_2 removal - 94.2%

	Consumption Rate	Lbs/lb Coal Fired	Lbs/lb SO_2 Absorbed
Makeup Materials			
Process Water	369 gpm	---	---
Lime (92.5 wt.% $Ca(OH)_2$, dry basis)	460 lbs/min	0.111	1.25
Soda Ash	13.7 lbs/min	0.003	0.037
Cake Production			
Dry Basis	804 lbs/min	0.194	2.18
Wet Basis	1,246 lbs/min	0.300	3.39

TABLE IV-7

OVERALL WATER BALANCE AT DESIGN CONDITIONS

	Inlet (gpm)	Outlet (gpm)
Process Water		
• Lime Slurry - Free Water	128.9	
- Chemically Combined	12.4	
• Filter Cake Wash	90.7	
• Pump Seal	20.0	
• Instrument Purge	0.5	
• Rain	0.6	
• Make-up (to Thickener Hold Tank)	116.2	
• Evaporation - Absorber		278.1
- Open Tanks		32.0
• Filter Cake - Free Water		53.1
- Chemically Combined		6.1
	369.3	369.3
Non-Contact Water		
• Fan Bearing Cooling	175	
• Vacuum Pump Seal	40	
• Fan Bearing Cooling		175
• Vacuum Pump Seal		40
	215	215

there will be a significant net evaporation of water from the system. The annual average evaporation rate from the thickener and hold tank is estimated to be 32 gpm and the average gain from rainfall is estimated to be about 0.6 gpm.

Since on the average water leaving the system will exceed the uses of water, a net amount of makeup water must be supplied to maintain the system water balance. This water is added to the thickener hold tank as required to maintain a constant liquid level. When the boiler is fueled with 5% S coal and operated at full capacity (300 Mw), the average rate of addition of makeup water is estimated to be 116 gpm.

2. Detailed Material Balance

A detailed material balance for design conditions is given in Tables IV-8 and IV-9. These tables present the material flows for all the streams according to the process flow diagram in CEA drawings Nos. 040044-1-1 and 040044-1-2. It should be noted that the water balance shown in Tables IV-8 and IV-9 has been simplified to reflect only process-related factors. Atmospheric effects (i.e., rainfall and evaporation from tanks), pump seal water and instrument purge water are not included in these tables; these effects are, however, considered in determining the net hold tank requirement given in Table IV-7.

The total flue gas at the intake of the FGD booster fans is 1,066,000 acfm at 300°F. This gas is roughly equally distributed between each of the two scrubber modules. The total saturated gas flow rate from both absorbers is 873,000 acfm at 126°F. Combustion gas from the oil-fired reheater is mixed with the saturated gas to reheat the exhaust gas 50F°. The total flow of reheated gas from both scrubber modules is 974,000 acfm at 176°F.

At design load, regenerated scrubbing solution is fed to each absorber at a rate of 1,840 gpm which is equivalent to about 4.2 gals/1000 acf of saturated gas. The combined flow rate to the top trays of scrubbing solution feed forward and absorber bottoms recycle is 2,710 gpm per absorber. This is equivalent to a liquid loading of 6.2 gals/1000 acf of saturated gas. The absorber bottoms are also recirculated to the bottom tray underside sprays at a rate of 1,600 gpm for each absorber. This spray rate corresponds to 3.7 gals/1000 acf of saturated gas. A bleed stream of 1,700 gpm of spend scrubbing solution is directed from each absorber to the corresponding reactor train. The difference between the feed forward to the absorbers and the bleed stream to the reactors, 140 gpm per absorber, reflects the evaporation of water to the flue gas. The total evaporation rate of both absorbers of 280 gpm is roughly equivalent to 1 gpm/megawatt.

At excess air levels (equivalent to 5-6% oxygen in the flue gas dry basis) normally encountered in utility power plant operations burning medium or high sulfur coal (coal containing 2.5% sulfur or higher), the level of oxidation is expected to be on the order of 5-10% of the sulfur dioxide removed. This material balance is based on a conservative estimate of the

TABLE IV-8

MATERIAL BALANCE - ABSORBER SECTION*

Stream No.	I.D. Fan Outlet 1	Absorber Inlet 2	Absorber Outlet 3	Combustion Air 4	Exit Flue Gas 5
Volume, ACFM	532,907	525,880	436,516	11,000	487,215
Temperature °F	300	303	126	60	176
Pressure, inches WG	+2.0	+11.5	+3.5	AMB	+2.0
Dry Gas, #/min	26,699	26,699	26,515		
H_2O Vapor, #/min	1,316	1,316	2,474		
Total Gas, #/min	28,015	28,015	28,989	850	27,839
SO_2, #/min	194.95	194.95	11.25		11.25
SO_2,PPM (Dry Vol)	3,471	3,471	200		200
Particulates, #/min	2.48	2.48	2.48		2.48

Stream No.	Combined Recycle 11	Spray Recycle 12	Tray Recycle 13	Absorber Bleed 14	Feed Forward 15	Tray Feed 16	Emergency Spray Water 31	Reheater Oil 32
H_2O, #/min	19,933.5	12,919.5	7,014.0	13,642.8	14,828.1	21,867.0	4,582.0	
Na_2CO_3, #/min					6.8			
NaOH, #/min					50.0			
Na_2SO_3, #/min	174.9	113.4	61.5	119.7	347.2	582.0		
$NaHSO_3$, #/min	666.9	432.2	234.7	456.4		95.0		
Na_2SO_4, #/min	1,542.9	1,000.0	542.9	1,056.0	1,016.3	1,559.2		
$Ca(OH)_2$, #/min								
$CaSO_3 \cdot 1/2\ H_2O$, #/min								
$CaSO_4 \cdot 1/2\ H_2O$, #/min								
$CaCO_3$, #/min								
Inerts, #/min								
CaCl, #/min	445.8	288.9	156.9	305.1	303.6	460.5		
Total, #/min	22,764.0	14,754.0	8,010.0	15,580.0	16,552.0	24,563.7	4,582.0	
% Solids, #/min	0	0	0	0	0	0	0	
Flow, GPM	2,481.0	1,608.0	873.0	1,698.0	1,837.0	27,101.0	550.0**	3.0
pH	4-8	4-8	4-8	4-8	7-13	6-11		
Temp, °F	120-140	120-140	120-140	120-140	120-140	120-140		AMB

* For scrubber module

** Emergency only, Max. Flow

Combustion Equipment Assoc.
Material Balance
for
Drawing 040044-1-1 Rev G
See Page 25

TABLE IV-9

MATERIAL BALANCE - REACTOR, SOLIDS DEWATERING AND RAW MATERIALS SECTIONS

	Absorber Bleed	Feed Forward	Lime Slurry Pumped	Reactor Lime Feed	Reactor Effluent	Filter Feed	Thickener Overflow	Cake Wash Water
Stream No.	14*	15*	17	18*	19*	20	21	22
H_2O, #/min	13,642.8	14,828.1	2,150.8	537.7	14,233.0	2,103.5	29,182.8	756.0
Na_2CO_3, #/min		6.8					13.9	
NaOH, #/min		50.0			50.2	7.3	101.3	
Na_2SO_3, #/min	119.7	347.2			349.4	51.0	704.9	
$NaHSO_3$, #/min	456.4							
Na_2SO_4, #/min	1,056.0	1,016.3			1,019.8	143.7	2,057.6	
$Ca(OH)_2$, #/min			852.0	213.0				
$CaSO_3 \cdot 1/2\ H_2O$, #/min					331.0	662.0		
$CaSO_4 \cdot 1/2\ H_2O$, #/min					39.4	78.8		
$CaCO_3$, #/min					5.8	11.6		
Inerts, #/min			69.2	17.3	17.3	34.6		
NaCl, #/min	305.1	303.6			305.1	44.5	615.5	
Total, #/min	15,580.0	16,552.0	3,072.0	768.0	16,351.0	3,147.0	32,676.0	756.0
% Solids	0	0	30.0	30.0	2.4	25.0	0	0
Flow, GPM	1,698.0	1,837.0	305.2	76.3	1,785.0	302.0	3,618.0	91.0
pH			10-14	7.0	6-13	7-13	7-13	
Temp, °F					120-140	120-140	120-140	AMB

	Cake	Filtrate	Soda Ash Solution	Soda Ash Feed	Soda Ash Solution Liquor	Make-up Water	Lime Slurry Recycle	Underflow Slurry Dilution Liquor	River Water (Startup)
Stream No.	23	24	25	26	27	28	29	30	33
H_2O, #/min	442.1	2,421.4	403.9		403.9	878.0	1,075.4		
Na_2CO_3, #/min			13.9	13.7	0.2				
NaOH, #/min	0.50	6.8	1.4		1.4				
Na_2SO_3, #/min	3.4	47.5	9.6		9.6				
$NaHSO_3$, #/min									
Na_2SO_4, #/min	10.0	138.8	27.9		27.9				
$Ca(OH)_2$, #/min							426.0		
$CaSO_3 \cdot 1/2\ H_2O$, #/min	662.0								
$CaSO_4 \cdot 1/2\ H_2O$, #/min	78.8								
$CaCO_3$, #/min	11.6								
Inerts, #/min	34.6						34.6		
NaCl, #/min	3.00	41.5	8.3		8.3				
Total, #/min	1,246.0	2,656.0	465.0	13.7	451.3	878.0	1,536.0		
% Solids	63.1	0	0				30.0		
Flow, GPM		299.0	50.0		50.0	105.0**	152.6	250.0***	135.0 max
pH		7-13	8-14		7-13			7-13	
Temp °F		120-140	120-140		120-140	AMB		120-140	

* Per module

** Includes water entering process from pump seals

*** Intermittent Maximum

Combustion Equipment Assoc.

Material Balance
for
Drawing 040044-1-2 Rev G

See Page 26

rate of oxidation of 10% of the SO_2 removed. The sulfite and sulfate concentrations indicated in Table IV-5 are consistent with the level required to precipitate sulfate as a calcium salt in balance with 10% oxidation rate (taking into account losses of Na_2SO_4 in the washed cake).

The total active sodium concentration in the feed forward to the absorbers is set at 0.45 M ($[Na_2SO_3]$ = 0.18 M, [NaOH] = 0.08 M, and $[Na_2CO_3]$ = 0.005 M). At a 10% level of oxidation and a high wash ratio, this results in an estimated Na_2SO_4 concentration of 0.47 M. The NaCl concentration estimated in the feed forward liquor is 0.34 M. This represents the steady-state level at which the rate of NaCl loss in the washed cake is equivalent to the chloride absorbed from the gas (1.8 lbs/min).

The feed forward rate of the regenerated liquor from the thickener hold tank to the absorber is controlled by the pH of the absorber bleed. At design conditions the absorber bleed pH is assumed to be 6. At this pH the absorber bleed will contain 0.31 M $NaHSO_3$ and 0.07 M Na_2SO_3. The Na_2SO_4 concentration in the bleed is increased to 0.52 M. The changes in the concentrations in the outlet and the inlet streams reflect 10% oxidation, absorption of 368 lbs/min of the SO_2 (both absorbers), and changes in the stream volumetric flow rate due to the evaporation of water in the absorbers.

The design feed rate to the two primary reactors is about 3,400 gpm of spent absorber liquor (1,700 gpm to each reactor). Lime slurry containing 30% solids is also fed to the reactors at a rate of 152 gpm (76 gpm to each reactor). The calcium hydroxide available in the solids is taken as 92.5%. Thus, the total feed of available calcium hydroxide to the two primary reactors is 426 lbs/min. This rate is equivalent to 1.00 moles of calcium hydroxide per mole of SO_2 removed in the absorbers.

The liquor from the primary reactor overflows into the secondary reactor where regeneration of scrubbing solutions is completed.

The composition of the liquor in the secondary reactor is dependent on the degree of regeneration. For design purposes, the secondary reactor is assumed to operate at a pH of 12. At this pH the regeneration reactions are carried beyond neutralization and some sodium hydroxide is formed. The composition of the secondary reactor liquor is: [NaOH] = 0.085 M, $[Na_2SO_3]$ = 0.19 M, and $[Na_2SO_4]$ = 0.48 M. The soluble calcium concentration in the liquor is less than 100 ppm. The liquor also contains sulfite/sulfate solids, inerts and a small amount of unreacted lime. The overall solids concentration in the slurry is 2.3%. The composition of the solids is estimated to be approximately 1% unreacted lime, 5% inerts (from the carbide lime slurry), 94% mixed calcium sulfite/sulfate solids (about 84% as calcium sulfite and 10% as calcium sulfate). The composition of these insoluble solids is essentially the composition of the insoluble solids in the waste cake.

The liquor from the two secondary reactors is pumped at a combined rate of 3,570 gpm to the thickener center well where the solids are allowed to settle. The underflow slurry is recycled around the thickener and a bleed from this recirculation loop is sent to the filters. The concentration of solids in the underflow slurry from the thickener is controlled at about 20-25% solids.

The filters are operated on overflow with the slurry level in the filter tubs controlled by the position of the overflow weir. At design load, the total slurry feed to the filters is 300 gpm and contains approximately 2,500 lbs of insoluble solids/hour (assuming 20% insoluble solids). About 60% of the total slurry fed to the filters is returned in the overflow to the thickener. The solids are filtered, forming a cake which is washed with water to remove sodium salts in the liquor entrained in the cake. The design wash rate is 90 gpm, which corresponds to a wash ratio of about 2.0 (volume of wash water/volume of entrained liquor). The combined filtrate (wash water and recovered liquor) is returned to the thickener. The washed cake is discharged at a rate of 1,246 lbs/min from the filter drum. The washed filter cake is estimated to contain 63 wt.% insoluble solids and about 1.3 wt.% soluble salts.

The soluble salts amount to 2.1% of the total solids present. NaCl accounts for about 40% of these solubles.

Clarified, regenerated liquor (including filtrate and wash water) overflows the thickener at the rate of about 3,600 gpm to the thickener hold tank from which it is pumped to the absorbers. Clarified liquor is also used to make up soda ash solution.

In this material balance, soda ash solution is shown as fed to the thickener, although the capability exists for feeding soda ash solution directly to the absorbers. The rate of soda ash makeup required to replace the sodium value lost in the cake is 13.7 lbs/min, equivalent to 2.3% of the SO_2 absorbed on a molar basis.

Based upon the total soda ash and lime feed rates, the overall system alkali stoichiometry is estimated to be 1.02 (moles of available CaO + Na_2CO_3)/(mole of $\Delta SO_2 + \Delta Cl_2$).

F. OPERATING AND CONTROL PHILOSOPHY

The development of the operating guidelines for the dual alkali system has been based on the minimization of operator interface to control and operate the system under varying conditions, mainly boiler load and sulfur content of the coal.

Detailed piping and instrumentation diagrams* (P&ID's) are given in CEA drawings 040044-1-1,3,4,5, and 6. It should be noted that instrument

*A set of large size CEA drawings may be obtained from the IERL Project Officer at the Environmental Protection Agency, Research Triangle Park, North Carolina. Because of the level of details in these drawings, they cannot be reduced, and therefore they are not included here.

redundancy or a backup system has been provided for all principal control loops. This includes fan controls, tank level controllers and indicators, and process liquor pH controllers.

The basic control philosophy for the four major process sections--absorber, reactor, solids dewatering, and raw materials handling--is discussed below.

1. Absorbers

The operation and control of the absorber section can be conveniently divided as it applies to the gas streams and the process liquors.

The control philosophy for this section of the system is summarized in Table IV-10.

a. Gas

Flue Gas Flow

The control of flue gas flow to the absorbers is based upon maintaining balanced pressure at the boiler I.D. fans. The control parameter therefore is the discharge pressure of the I.D. fans. A pressure indicating controller maintains this pressure in the range of -0.5 to 1.0 inches W.G. (depending upon boiler load) by adjusting the speed of the dual alkali system booster fans. One pressure controller is used. The signal is sent to each booster fan speed controller through a precalibrated bias. The bias adjusts the signal/response to account for mechanical differences in the two booster fan fluid drive units and effect equal speeds on both fans and parallel fan tracking of boiler load.

Normally both absorbers and fans are operated together with each train taking half the boiler load. However, a common duct between the booster fan discharges and appropriate dampers provide the flexibility of operating only one absorber at low boiler load (<60% load). This allows for maintenance on one absorber module while the system remains in service.

Flue Gas Reheat

Flue gas reheat is provided through injection of heated air/gas into the saturated absorber discharge. The heated air/gas is generated by combining ambient air with combustion gas from a No. 2 oil-fired burner. The operation of the burner is controlled based upon the degree of flue gas reheat as measured by the differential temperature between the reheated gas and saturated absorber discharge. The differential temperature controls both the burner oil and combustion air fans to provide the preset degree of reheat (usually 50F°).

The dilution air which is mixed with the burner combustion gases is normally maintained at a constant rate regardless of load.

TABLE IV-10

CONTROL PHILOSOPHY FOR THE ABSORBER SECTION

Controlled Variable	Approximate Variable Range (per absorber)	Controlled By	Control Function	Control Parameter	Parameter Range	Control Tag
Gas:						
Flue Gas Flow	106,000-533,000 acfm	Fan Speed	Gas Flow Balance	Fan Suction Pressure	-1.0-0.5" W.G.	DF-101/DF-102
Reheater Oil Flow	0-3.0 gpm	CV-56	Reheat Scrubbed Gas	Gas Differential Temperature	0-50°F	TIC-1/TIC-2
Process Liquor:						
Feed Forward Flow	280-1850 gpm	CV-3/CV-26	SO_2 Removal	Bleed Liquor pH		FFIC-1/FFIC-6
Soda Ash Feed Flow	0-100 gpm*	CV-60/CV-61	Sodium Makeup	Flow	0-100 gpm*	Manual (FIC-2)
Tray Recycle Flow	260-870 gpm	CV-2/CV-25	Top Tray pH Tray Loading at Low Flow	Flow	260-870 gpm	Manual (FIC-1/FIC-3)
Spray Recycle Flow	1600 gpm	---	Gas Quench	Manual	---	Manual
Absorber Bleed Flow	260-1700 gpm	CV-4/CV-27	Flow Balance	Absorber Tank Level	4'-5'	LIC-1/LIC-6

*Total Flow

Emergency Water Sprays

Emergency water sprays are provided on the inlet ducting to each absorber to protect the absorber linings from temperature excursions. These emergency sprays are designed to quench the incoming flue gas in the event of a failure of the spray system. The sprays are activated either by a high temperature reading in the absorber discharge gas or low flow in the spray recycle line.

b. Process liquor

The removal of SO_2 from the flue gas is the basic function of the absorber. Consequently, the principal control parameter for operation of the absorbers is the amount of SO_2 removed. Since the pH of the bleed liquor can be accurately related to the amount of SO_2 removed, it can be used to provide this control. Additional control parameters for the absorbers include the flow of the recycle streams as well as the bleed stream.

Absorber Bleed pH/Feed Forward Flow Rate

The feed forward of regenerated alkaline, scrubbing solution to the absorber is determined by the pH of the absorber bleed liquor. The flow is adjusted to maintain the bleed liquor pH within the range prescribed by the required SO_2 concentration in the exit flue gas. Initially the bleed pH will be controlled to a value of about 6, which is estimated will result in an SO_2 concentration in the exit gas of less than 200 ppm. The actual correlation between pH and SO_2 removal will be determined during the early periods of operation and the actual set points modified accordingly. The option exists for operating the absorber feed forward rate on flow control with manual adjustment of the flow set point. This may be required during periods of maintenance on the bleed liquor pH monitors.

Soda Ash Makeup

An additional stream that feeds the absorbers is the soda ash makeup solution. This solution, at a constant flow rate (usually 50 gpm), is directed to either the thickener or the absorbers. The principal mode of operation is to feed soda ash solution to the absorbers. If both absorbers are in operation, it is distributed equally between them. While the total flow of soda ash solution is maintained on flow control, the distribution of solution to either or both absorbers is manually adjusted. Soda ash flow can be switched to the thickener during periods of abnormally high soda ash makeup rates, which would affect the bleed liquor pH. Further details on the control of the preparation of the soda ash makeup solution are given in the discussion of raw materials handling.

Tray Recycle

The tray recycle flow rate is normally controlled at 650 gpm (range of 260-870 gpm). Given the varying feed forward rate, the tray recycle is needed to insure a proper liquid loading of the trays. It also reduces the pH of the total liquor fed to the tray, thus minimizing CO_2 absorption and the potential for $CaCO_3$ precipitation. The pH of the tray feed (combined tray recycle and feed forward) will normally fluctuate within the range of 7.5-10.5. The set point for the tray recycle flow rate may be changed to maintain the pH within this range; however, such adjustments would normally be required only if the pH falls outside the range for extended periods of time (a few hours or more). Hence, the pH of the combined liquor feed to the top tray is monitored and alarmed, but is not used for automatic control of the recycle flow. The flow is maintained in flow control and is adjusted manually as required. Increasing the recycle flow rate will lower the pH of the combined stream; decreasing the recycle flow rate will have the opposite effect.

Spray Recycle

The spray recycle plays the important role of quenching the incoming flue gas and protecting the absorber lining. The sprays are operated at a constant flow rate of 1,600 gpm independent of boiler load. The flow rate is manually controlled when required by the appropriate adjustment of the spray recycle valve position.

Absorber Bleed Stream

The flow rate of the bleed stream is controlled by the liquid level in the absorber tank, which is normally maintained at a height of 4.5 ft (range 4-5 ft). The flow rate of the bleed stream is expected to vary within 260-1,700 gpm over the full range of boiler load and coal sulfur content. The option exists for operating the absorber bleed rate on flow control with manual adjustment of the flow set point. This would be required during periods of maintenance on the absorber level control loop.

2. Reactor Section

The optimal operation of the reactor system involves controlling the extent of regeneration scrubbing solution to maximize lime utilization and produce waste solids with good dewatering characteristics. The pH of the secondary reactor bleed stream is a measure of the extent of regeneration and thus becomes the controlling factor in this step. A summary of the control philosophy for the reactor section is given in Table IV-11.

TABLE IV-11

CONTROL PHILOSOPHY FOR THE REACTOR SECTION

Control Variable	Variable Range (per reactor train)	Controlled By	Control Function	Control Parameter	Parameter Range	Control Tag
Primary Reactor Bleed Flow	270-1800 gpm	Overflow	Reaction Time/ Flow Balance	Tank Level	12 ft	Overflow
Secondary Reactor Bleed Flow	270-1800 gpm	CV-11/CV-14 or Overflow	Reaction Time/ Flow Balance	Tank Level	18-30 ft	LIC-2/LIC-3 or Overflow
Lime Feed Rate	30-120 gpm	CV-12/CV-13	Solid Waste Properties/Lime Utilization	Second Reactor Bleed Liquor pH	11-12.5 ft	AIC-1/AIC-2

a. Individual Reactor Controls

Tank Level/Effluent Flow - Primary Reactors

The absorber bleed streams are mixed with carbide lime in the primary reactors and subsequently overflow into the secondary reactors. Thus, the tank level in the primary reactor is constant at the overflow level, and the effluent flow rate is the same as the combined feed of absorber bleed and carbide lime slurry. The hold-up time in these reactors is dictated by the flow rate of the combined feed streams. The normal operating range for the hold-up time is 3-10 minutes.

Tank Level/Effluent Flow - Secondary Reactors

Each of the secondary reactors is provided with one reactor transfer pump and an overflow line. Thus, the effluent (or bleed) from the secondary reactors can be fed to the thickener either by pumping it on level control or by allowing the tank to operate on overflow. In either case, the secondary reactors are normally operated at hold-up times greater than 30 minutes.

Under typical conditions of boiler load and coal sulfur content, it is expected that the secondary reactors will be operated on overflow (level of 30 ft). The flexibility is also provided to operate the secondary reactors at lower liquid levels by using the reactor transfer pumps (with flow controlled on liquid level). This allows adjustment of secondary reactor hold-up times if desired during sustained periods of extremely low boiler loads. The pumps are also used to drain the reactors if they are to be taken out of service for maintenance.

Bleed pH/Lime Slurry Feed

The amount of lime fed to the primary reactor of each train is controlled by the pH of the corresponding secondary reactor bleed stream. The bleed pH is automatically controlled within the range of 11-12.5 (initial set point of 12) by adding more lime to increase the pH or reducing the lime feed to decrease it.

There are occasions, however, when it is desirable to override pH control to prevent overfeeding of lime. This could occur, for example, whenever there is a rapid and significant change in boiler load or coal sulfur content, or the pH of the secondary reactor experiences a rapid drop due to some upset condition. Under such circumstances, it is very likely that lime will be overfed to the primary reactors for some period of time since the lag time for pH sensors to detect the effect of lime addition is greater than one-half hour (primary plus secondary reactor hold-up times).

To avoid this problem a maximum lime feed rate controller has been provided. It sets a limit on the lime feed rate which is based in direct proportion to the absorber bleed stream flow. Thus, the secondary reactor bleed pH controls the lime feed rate as long as this feed rate does not exceed the maximum rate established by the overriding controller. In the event that only one reactor train is in operation but both absorbers are operating, the maximum lime feed rate to the single reactor train is proportional to the sum of both absorber streams. This is accomplished automatically when placing a single reactor train in service.

b. One vs. Two Reactor Train Operation

The number of reactor trains in operation will generally be dictated by the boiler load and the sulfur content of the coal. Normally, with typical coal (~3.5% S) only one reactor train is required with the secondary reactor usually operated on overflow.

It is highly desirable, of course, to minimize switching between one and two reactor train operation. Momentary upsets or short-term changes in boiler loads therefore are not considered as a basis for changing the number of trains in operation. However, during sustained periods of operation (more than three to four hours) with the boiler firing coal having a sulfur content of greater than about 4.0% (>3000 ppm SO_2 in flue gas - dry basis) operation of both reactor trains would be desirable unless the boiler load is consistently low (less than about 70%). Short-term low load conditions (either with one or two trains in service) can usually be accommodated if necessary by simply switching from operation of the secondary reactor in overflow mode to level control (and possibly readjusting the pH setpoint).

As indicated in the discussion of lime feed rate control, whenever the number of reactor trains is changed, the absorber bleed flow totalizer is automatically activated accordingly.

3. Solids Separation

The separation of the solid wastes from the regenerated scrubbing solution is accomplished in a two step operation: thickening of the slurry followed by vacuum filtration. A summary of the control philosophy for the solids separation section is given in Table IV-12.

a. Slurry Thickening

Thickener

The reactor effluent fed to the thickener contains 2-5% solids. The thickener is generally operated to provide a slurry underflow containing 20-25% solids. This solids concentration provides for reasonable control of the filter operation. The solids generated in the reactor system settle very well and will normally thicken to a much higher solids level, up to 40% or more, if not controlled. Hence, clarified liquor from the

TABLE IV-12

CONTROL PHILOSOPHY FOR THE SOLIDS SEPARATION SECTION

Slurry Thickening	Variable Range	Controlled By	Control Function	Control Parameter	Parameter Range	Control Tag
Thickener Underflow Recycle Flowrate	250-600 gpm	---	Prevent Solids Deposition			Manual
Dilution Liquor Flow Rate	0-260 gpm	CV-17	Suspended Solids Concentration	Underflow Specific Gravity	1.20-1.26	DIC-1
Makeup River Water Flow Rate	50-150 gpm	CV-24	Maintain System Water Balance	Hold Tank Level	8-16'	LIC-14
Dewatering						
Drum Speed	0.8-1.2 rpm		Filtration Rate	Cake Thickness	1/4-1/2"	Manual
Drum Submergence	8-10"	Weir position	Filtration Rate	Cake Thickness	1/4-1/2"	Manual
Wash Water Flow Rate	40-180 gpm*	---	Minimize Na^+ Losses			Manual
Filtrate Sump Recycle	50-700 gpm	CV-53	Flow Balance	Filter Sump Level	2.5-5'	LIC-9

*Per filter

hold tank is used to control the solids level in the thickener underflow that is pumped to the filters. A density sensing element is located at the discharge of the underflow pumps to measure the slurry concentration and adjust the rate of dilution liquor addition in order to maintain a specific gravity consistent with 20-25% solids. Part of the slurry is recirculated past the filters and returned to the solids zone in the thickener in order to maintain flow through the lines, thus preventing solids deposition. The underflow and recycle flow rates are manually set by the appropriate opening of the corresponding valves.

Thickener Hold Tank

The clarified liquor from the thickener overflows into the hold tank. Makeup river water, required to maintain the water balance in the system, is fed to the hold tank on level control. The hold tank liquor, which is regenerated scrubbing solution, is pumped to the absorbers. As previously indicated in the absorber section, the flow rate of the feed forward to each absorber is controlled by the pH of the absorber bleed stream.

b. Vacuum Filtration

Filter Units

The most important factors in assessing filter performance are the solids content and the level of sodium salts contained in the filter cake (i.e., solids dewatering and washing). These are primarily a function of the quality of the solids produced in the reactor system, the thickness of the filter cake and the amount of cake washing. In general, optimal performance is achieved with a cake thickness of about 3/8 inch (a range of 1/4-1/2 inch) and a wash ratio in the range of 1.5 to 3.0 (gals of wash water/gal of water in the filter cake). The wash water rate, therefore, would be a function of the rate of discharge of wet cake and the moisture (or solids) content of the cake. Since the solids content varies only slightly and the drum speed is usually fixed, the wash water flow rate can be preset as long as cake thickness is controlled in the proper range. If necessary, it can be readjusted based upon the weight of cake produced as measured by the conveyor belt weigh element. The wash flow rate to each filter is initially set at 120 gpm.

The cake thickness is controlled by four parameters, three of which are preset: drum speed (1.0 rpm), drum submergence (8-10 inch), and the filter bridge valve position (6 o'clock). The fourth parameter, the underflow slurry concentration, is used as the primary control parameter to maintain proper cake thickness. As indicated in the thickener section, the slurry concentration is kept at a suspended solids level of about 20-25%. However, when the slurry concentration deviates appreciably from this range and cannot be further adjusted by dilution liquor, operator interface is required to either put filters in service or take them off line.

Number of Filter Units in Service

The number of filters in operation is dictated by the thickener underflow slurry concentration and the amount of diluting hold tank liquor added. Whenever the dilution liquor has been automatically reduced to its minimum rate and the slurry concentration is below about 15%, a filter needs to be taken off line. Similarly, when the dilution liquor rate is at its maximum and the slurry concentration exceeds about 28%, a filter needs to be put in service.

Filtrate Sump

The filtrate is collected in filtrate receivers and directed to the filtrate sump from which it is pumped to the thickener. The filtrate sump is maintained at a constant level by controlling the flow in the sump recycle stream. The excess fluid goes to the thickener.

4. Raw Materials

The dual alkali process requires the addition to the system of two raw materials: carbide lime and soda ash makeup. A summary of the control philosophy for the raw materials handling and feed systems is given in Table IV-13.

a. Carbide Lime

Lime Feed to Day Tank

Carbide lime is available to LG&E as a slurry containing 25-30% insoluble solids. The lime for the dual alkali system is stored in a day tank which is filled batchwise as the slurry level drops below the preset point. Initially, the preset level corresponds to about 60% of the tank capacity and requires one or two fillings per shift. The preset level may be changed, however, and the contents of the lime tank may be allowed to drop as low as 25% of the tank capacity before proceeding to refill it.

Lime Tank Recycle

The amount of lime added to the primary reactors is controlled by the pH of the corresponding secondary reactors with override limitation on maximum feed based upon absorber bleed flow, as previously discussed. The lime slurry is recirculated around the lime feed tank to maintain sufficient flow through the pipes to prevent solids deposition. A bleed from this recycle is fed to the primary reactors. A control valve on the day tank return line is interconnected with the control valves on the lime slurry feed to the individual reactors. The recycle valve is used to maintain enough pressure in the feed lines to each primary reactor to ensure sufficient slurry feed and adequate feed control. The valve position on the slurry return line to the day tank is controlled by the larger of the flows to each of the primary reactors. As the larger of the reactor feed rates increases (decreases), the control valve on the return line closes (opens) to maintain the feed line pressure consistent with a reasonable range of operation for the feed control valves.

TABLE IV-13

CONTROL PHILOSOPHY FOR THE RAW MATERIALS SECTION

Control Variable	Variable Range	Controlled By	Control Function	Control Parameter	Parameter Range	Control Tag
Carbide Lime						
Lime Feed	Batch	CV-7	Sufficient Lime Inventory	Lime Tank Level	15'-21'	
Lime Recycle	150-285 gpm	CV-8	Prevent Solid Deposition/Enough Pressure Rxtr Feed	Primary Reactor Feed	150-285 gpm	VPC-1
Soda Ash						
Soda Ash Feed	40-80 tons every week	Batch	Sufficient Soda Ash Inventory	Soda Ash Silo Level	12'-30'	Manual
Hold Tank Liquor Flow	30-60 gpm	CV-22	Soda Ash Solution Liquor	Flow Rate	30-60 gpm	FIC-2
Soda Ash Solution Flow	30-60 gpm	CV-23	Flow Balance	Soda Ash Tank Level	4'-6'	LIC-5

b. Soda Ash

Soda Ash Silo

Dense soda ash is received and stored in the soda ash silo, which has a capacity of approximately 140 tons. Maintaining the silo inventory will normally require two to four 20-ton trucks every week, depending on the system demand.

Soda Ash Solution Tank

The soda ash is fed to the soda ash solution tank by means of a weigh feeder. The amount fed can be set manually and readjusted daily or weekly as required; or it can be controlled automatically in proportion to the amount of lime fed to the reactor system. (The amount of lime fed dictates the amount of waste cake produced and ultimately the rate of soda ash lost in the washed cake.) On automatic control, the bias in the controller would be adjusted periodically (about once a week) to maintain a constant thickener overflow liquor specific gravity. If the specific gravity is increasing, the soda ash should be decreased. Similarly, if the specific gravity is decreasing, the soda ash feed should be increased.

Soda ash solution is normally prepared using clarified liquor from the thickener hold tank. Solution liquor from the hold tank is pumped (from the discharge of the hold tank pumps) to the soda ash solution tank at a constant rate, nominally 50 gpm. The resulting soda ash solution is pumped to the absorbers or to the thickener on level control. This operation is described in more detail in the Absorber Section.

V. DUAL ALKALI PLANT CONFIGURATION AND EQUIPMENT

A. PLANT LAYOUT

Cane Run Unit No. 6, the last and largest of six boilers at the Cane Run Station, is located at the north end of the plant complex. A photographic overview of the north end of the plant in Figure V-1 shows the boiler/turbine house, the two parallel electrostatic precipitator sections and exhaust stack of Unit No. 6 and the dual alkali facility. The two SO_2 absorbers have been installed behind the precipitators one on each side of the stack. The chemical plant consisting of the reactor system, dewatering equipment, and raw materials preparation areas are sited north of the boiler and scrubbers. Figure V-2 shows more detail on the gas ducting and reheaters.

The building shown behind the thickener houses the filters and associated filtration equipment; the reactor, lime feed, and hold tank pumps, and the system sumps. The two reactor trains are located next to and are accessed from the filter building. The soda ash silo and soda ash solution tank are located in the north-east corner of the facility, adjacent to the thickener in the foreground of Figure V-1. The thickener overflow tank is located between the thickener and the filter building; and the lime feed day tank is located behind the filter building.

The relative location of the reactor trains and the lime slurry tank is shown in more detail in Figure V-3. The elevation of the primary reactors, secondary reactors, and thickener is designed to allow operation of the regeneration system completely in an overflow mode.

The general plot plan for the dual alkali system is given in Drawing No. 040044-2-0[a]. General arrangements of the various parts of the system are given in the following drawings: absorbers, lower and upper plans--040044-2-1/040044-2-4; thickener, hold tank, and soda ash silo plan--040044-2-5; reactors and vacuum filter building plan--040044-2-6; absorber sections--040044-2-8/040044-2-9; thickener hold tank and silo sections--040044-2-10; and reactors and vacuum filter building sections--040044-2-11.

B. PROCESS EQUIPMENT

The following is a brief description of the major pieces of equipment. Further details and specifications can be found in the appendix.

---

[a]A set of large size CEA drawings may be obtained from the IERL Project Officer at the Environmental Protection Agency, Research Triangle Park, North Carolina. Because of too many details, these drawings cannot be reduced; and therefore, are not included here.

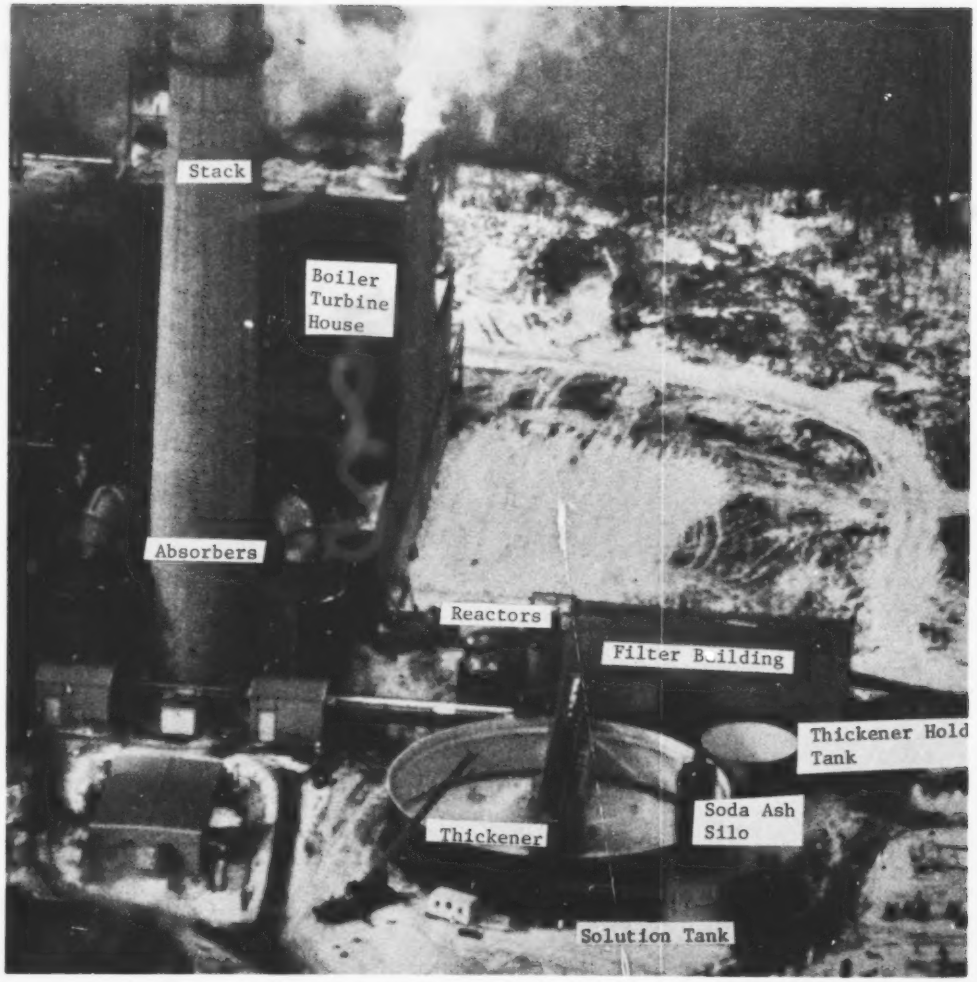


Figure V-1: Overall View of Cane Run Unit No. 6 and the Dual Alkali System

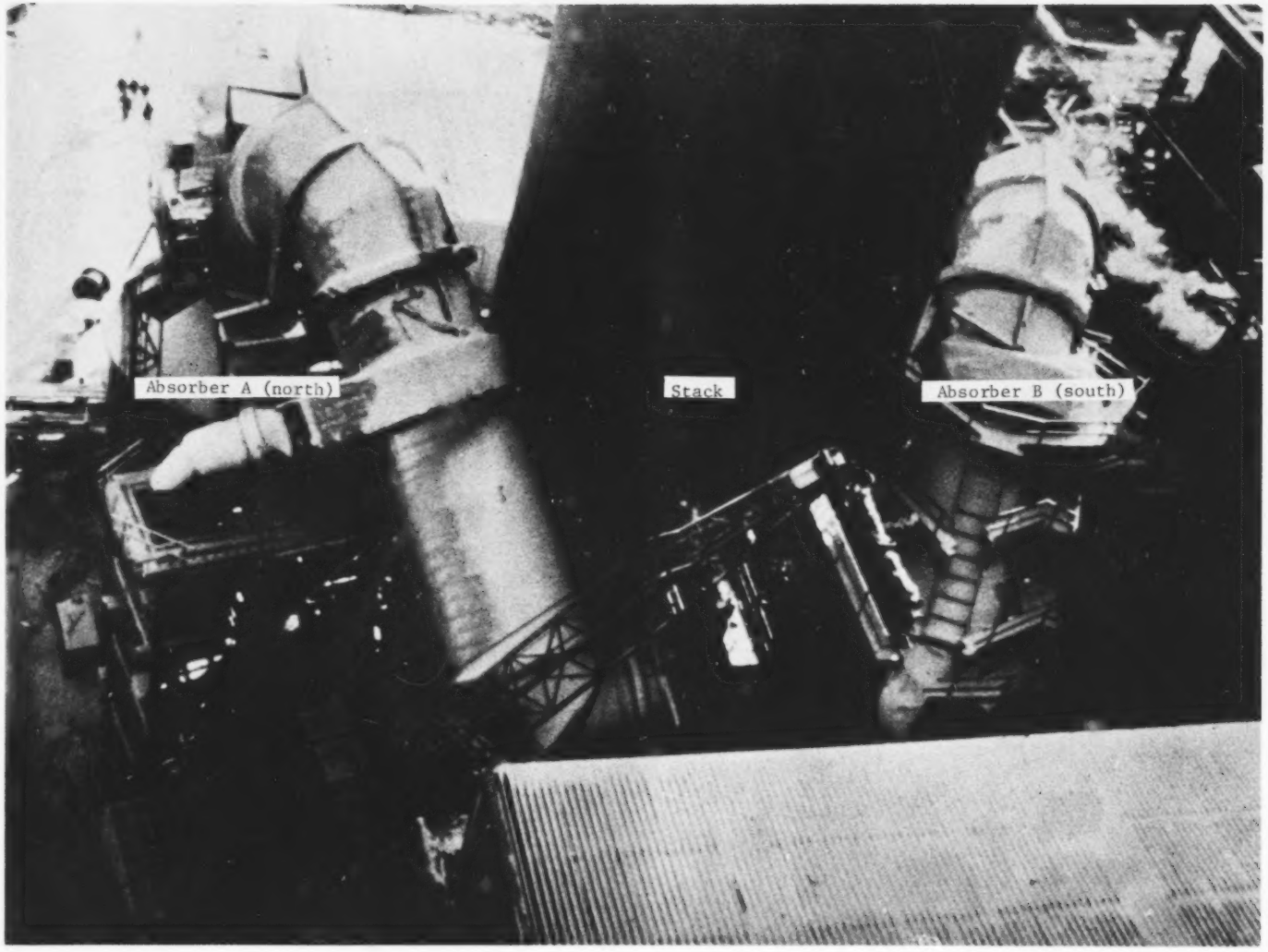


Figure V-2: Gas Reheaters and Absorber Ducting

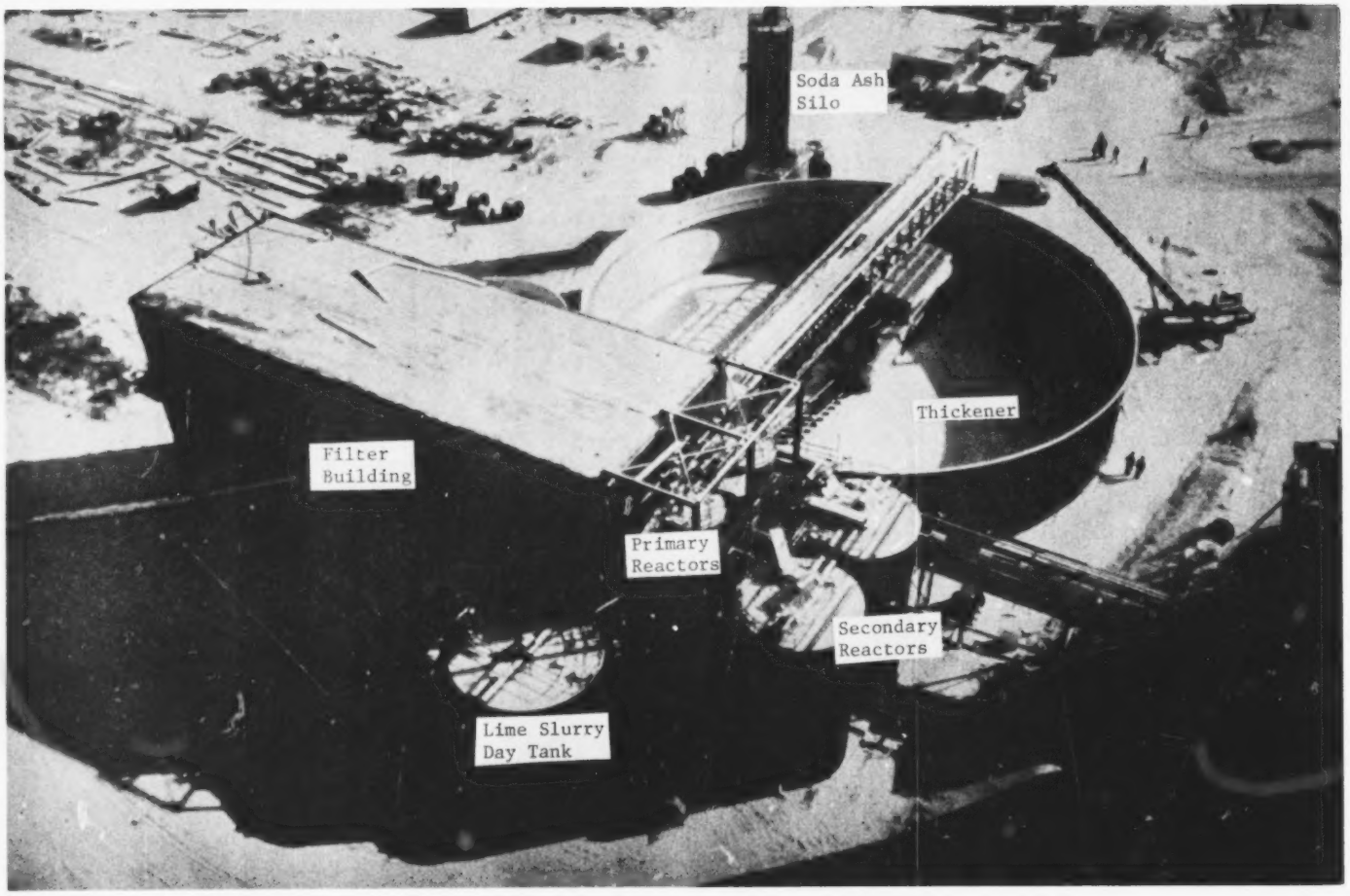


Figure V-3: Overall View of the Chemical Plant

1. Absorbers

Each absorber is a carbon steel vessel, cylindrical in shape and with a conical head. The overall height of the absorbers including the overhead discharge duct is 75 ft. The absorbers themselves are 55 ft high and 32 ft in diameter. The vessels are internally coated with reinforced polyester lining.

The internals of each absorber consist of an array of spray nozzles, two trays, and one demister. The sprays beneath the bottom trays are constructed of fiberglass headers with 316 stainless steel nozzles. The trays in each absorber are fabricated out of 317L stainless steel. At design conditions these trays have a liquid loading of 4.0 gallons/1000 acf. The pressure drop across these trays is 4-6 inches W.G. and the overall gas pressure drop through the dual alkali system is less than 9.5 inches W.G. The four pass, chevron type, mist eliminators are made of noryl.

An 11 ft 8 in carbon steel duct takes the flue gas discharged by each of the two parallel boiler induced draft (I.D.) fans and directs it to the corresponding booster fan. Each booster fan is fabricated out of A441 steel, and is driven by a 1250 HP motor with fluid drive speed control. The gas exits the fans and enters the absorbers through a 13 ft 10 in duct. The ducts to the two booster fans are interconnected by the by-pass duct. Five dampers provide flexibility in the operation of the absorbers. Each booster fan inlet has an A36 steel, guillotine type, damper to isolate the corresponding absorber while the flue gas is desulfurized by the other absorber. In addition the flue gas may by-pass entirely the FGD system by opening the multi-louver 316L stainless steel by-pass damper while keeping the booster fan and absorber outlet duct dampers closed. Each outlet duct from the absorbers 13 ft in diameter, has a 316L stainless steel, guillotine type, damper which is always maintained at the same position, open or closed, as the corresponding booster fan inlet damper. All the ductwork carrying saturated flue gas from the scrubber to the reheater is made of carbon steel, internally coated with reinforced polyester lining. The ductwork carrying reheated flue gas from the reheater to the stack is made of 317L stainless steel.

The continuous recycle flow through the sprays and trays, essential for operation of the absorbers, is ensured by a back-up pump provided for each absorber. Provisions have also been made to bring the back-up pump on-line, in case of failure of the primary pump without shutting the absorber down. The pumps are made of rubber lined cast iron and the process piping is made of fiber reinforced plastic (FRP).

2. Reactors

The primary reactors are cylindrical tanks, 11 ft in diameter and 14 ft in height, constructed of 316L stainless steel and equipped with a pitched blade turbine agitator driven by a 7.5 HP motor. The primary reactors are top fed from a 12-in line for the absorber feed and a 3-in line for the lime reactant slurry.

The overflow conduits from the primary to the secondary reactors have been specially designed. Each conduit is a trough, trapezoidal in cross section, with a removable top which allows easy access to the duct for cleaning purposes, if required.

The secondary reactors are also cylindrical tanks 20 ft in diameter and 33 ft in height. The tanks are constructed of carbon steel lined with glass reinforced polyester. Each reactor is equipped with a pitched-blade turbine agitator driven by a 25 HP motor and a transfer pump to direct the slurry product to the thickener. The transfer pumps are made of rubber lined cast iron. A 24-in overflow port that feeds directly to the thickener is also available for operation of the reactors in an overflow mode.

3. Solids Dewatering

The thickener is a tank 125 ft in diameter and 23 ft in height with a carbon steel shell and concrete bottom. Both shell and bottom are lined with glass reinforced polyester. The outer wall of the thickener extends 3 ft above the overflow weir providing a surge capacity equivalent to the capacity of all other tanks in the system to allow storage of the total system liquor and to provide capacity for temporary water balance upsets.

The solids at the bottom of the thickener are swept by a rake driven by a 5 HP motor. The position of the rake is adjusted for the accumulation of solids at the bottom by a 3 HP rake lifting motor. The center feed well extends half-way down the tank and is 12 ft in diameter. Both the rake and the center feed well are made of rubber covered carbon steel.

The clarified liquor overflows through a 30-in line to the thickener hold tank. The hold tank is a cylindrical vessel 36 ft in diameter and 23 ft in height made of carbon steel and lined with glass reinforced polyester. A rubber lined cast iron pump is used to transfer liquor from the hold tank to the absorber. This pump also supplies dilution liquor to the thickener underflow slurry as well as solution liquor to the soda ash make-up solution tank. A back-up pump is provided for this operation.

The thickener underflow slurry is pumped in a recirculation loop back to the thickener. (A back-up pump is provided.) A bleed from this recirculation loop is fed to the filters. Final dewatering of the waste cake takes place in the three vacuum filters. The filtering surface in these filters is a rotating drum 8 ft in diameter and 16 ft in length. A blow back fan and a rubber tipped 317 stainless steel scraper is used to discharge the cake. The slurry in the filter tub is agitated by a counter weighted rocker arm and the slurry level is controlled by an adjustable overflow weir. The drum and agitator are made of 317 stainless steel and the filter media is made of polypropylene. Soluble sodium salts are washed from the cake by a cake wash assembly consisting of two banks of drip wash nozzles and a drag net.

4. Raw Materials Preparation

This section of the system consists, basically, of three pieces of equipment, the day-tank for lime slurry, the soda ash silo, and the soda ash solution tank.

The lime slurry tank is a cylindrical vessel, 24.5 ft in diameter and 24.5 ft in height, made of carbon steel and equipped with a 20 HP agitator. The lime slurry is pumped in a recirculation loop back to the top of the lime slurry tank. (A back-up pump is provided.) The lime feed to the primary reactors is taken from this recirculation loop. Ni-hard alloy was used in the manufacture of these pumps.

The soda ash silo is a cylindrical tank, 12 ft in diameter and 34.5 ft in height, with a conical bottom, 6 ft bottom diameter and 5.5 ft in height, made of carbon steel. It is equipped with a cone vibrator, a baghouse at the top, as well as a pneumatic system to top-feed the silo. A weigh feeder is used to transfer the soda ash from the silo to the solution tank. A fan has also been installed to blow back the warm, humid vapors from the solution tank. Otherwise these vapors could cause the accumulation of soda ash crystals in the feed chute and weigh feeder thus interfering with the normal operation of the soda ash make-up solution.

The soda ash solution tank is a small cylindrical vessel (6 ft diameter, 8 ft height) made of carbon steel and lined with glass reinforced polyester. It is equipped with a 1.5 HP agitator.

C. OFFSITES AND AUXILIARIES

The offsites required for the dual alkali system at Cane Run Unit No. 6 include: services for electrical supply, water supply and instrument air; oil for reheating the wet flue gas, raw materials receiving and storage facilities; a wet chemical analytical laboratory; and shop facilities for repair and maintenance of machinery and instruments. Except for electrical service, all of these offsites existed and are available at Cane Run Station. An electrical substation including appropriate step-down transformers has been installed for the dual alkali system at Unit No. 6.

1. Electrical Power

Electrical power available at Cane Run Station for operation of the system include the following:

4,160 V ac, three-phase, 60 hertz

480 V ac, three-phase, 60 hertz

120 V ac, single-phase, 60 hertz

A feed line will be taken from the existing 14 KV substation and the voltage will be reduced in a step-down transformer to 4,160 volts. Step-down transformers for further voltage reduction to 480 volts and from 480 volts to 120 volts have been installed at LG&E.

The power requirement for the system is estimated at about 1% of the total power generated by the boiler at peak load. However, since the design gas flow rate for the dual alkali system at LG&E is higher than the maximum flow rate, the estimated electrical energy requirement is a conservative one. A summary of the ancillary requirements for the dual alkali system is given in Table V-1.

2. Water Supply

The maximum water requirement for the system is estimated at 450 gpm, not including the water associated with the slurried lime feed. Of the 450 gpm, approximately 230 gpm are required for process streams and about 220 gpm for non-contact cooling.

River water is used for all water requirements. The river water is available at the following conditions:

Water Supply Pressure	50-100 psig
Water Temperature	35-90°F
Total Dissolved Solids	300-500 ppm
Suspended Solids	50-500 ppm
pH	6-8
Na_2SO_4	20-200 ppm
Hardness	80-250 ppm
$CaCO_3$	50-250 ppm
Fe	0.1-30 ppm
Mn	0.15-2.5 ppm
NaCl	10-100 ppm

An in-line filter is used to filter the water supply to the system to prevent solids from entering the pump seals and filter spray nozzles.

TABLE V-1

ANCILLARY REQUIREMENT FOR THE DUAL ALKALI SYSTEM
LOUISVILLE GAS AND ELECTRIC
CANE RUN UNIT NO. 6

Design Capacity (300 Mw, 5% S)	
Electrical Power	3.1 Mw
River Water	450 gpm
No. 2 Fuel Oil	343 gal/hr
Typical Operation (180 Mw, 3.8% S)	
Electrical Power	1.7 Mw
River Water	370 gpm
No. 2 Fuel Oil	206 gal/hr

3. Instrument Air

Air is available at Cane Run Station at 60-125 psig. Air is used only for instruments and air-operated controls. The total amount of air used for the process is small and existing compressor capacity at the station is adequate to supply the air.

4. Oil

No. 2 fuel oil is used to reheat the wet exhaust gases from the absorbers. The oil requirement for the dual alkali system to provide 50F° of reheat at design load is estimated to be 343 gallons/hour.

5. Carbide Lime Facility

Carbide lime slurry will be used to regenerate the spent sodium solution. Since receiving, handling, and storage facilities for the carbide lime slurry already exist at the plant, only a day tank to store lime slurry for the process has been installed as part of the dual alkali system.

Carbide lime is available as a slurry containing 30% dry solids. The slurry is shipped in LG&E barges to the Cane Run Station. The slurry is then pumped from the barge to an agitated storage tank from which it is pumped to the dual alkali day tank as required.

In general, calcined lime, hydrated lime, or carbide lime may be used to regenerate the spent sodium solution. While carbide lime is cheaper than commercial lime, it is not available at most locations. Normally, calcined lime would be used. It would be slaked and fed to the system as a slurry, and therefore would not be considered an offsite. It is included as an offsite here because the carbide lime for facility already exists.

6. Laboratory and Shop Capabilities

The Cane Run Station has the necessary laboratory and shop capabilities for the maintenance and operation of the dual alkali system. No additional facilities are required. The equipment needed for wet chemical analyses is small and has been incorporated in the existing plant control laboratory.

a. Laboratory Capability

The following laboratory equipment and materials are needed for the chemical and physical testing required during the operation and testing period of the dual alkali system:

- Analytical balances.

- Atomic absorption spectrophotometer.

- pH meter with electrodes for standard pH measurements and lead electrodes for sulfate titration.

- Forced draft-type oven with temperature control ($\pm$ 0.5°C).

- Automatic burets and magnetic stirrers.

- Distilled water and various reagents for wet analyses.

- Assorted glassware for sample preparations.

Much of this equipment may already be part of an existing control laboratory at a power plant for use in monitoring and analysis of coal, cooling water, boiler feed water, and waste streams; or can be easily included as a part of the control laboratory equipment. In some cases, special analyses for metal ions (calcium, sodium, and magnesium) requiring the use of an atomic absorption spectrophotometer can be performed by outside testing laboratories.

b. Shop Capability

LG&E carries out their own plant construction. The Cane Run Station has adequate shop facilities to operate and maintain the boilers and the existing direct lime scrubbing systems. The shops are equipped with tools and equipment worth over $3.0 million, including 157-ton capacity crane.

In general, the required shop capacity for dual alkali systems includes the following:

- Crane capacity to lift motor, pump, valves, etc., which need occasional maintenance.

- Machine shop to machine relatively simple surfaces, thread pipes, etc.

- Welding equipment, both in shop and in field.

- Instrument shop to check out instruments.

- Electrical shop.

D. MECHANICAL TESTING OF EQUIPMENT

Mechanical testing of the dual alkali system began in December, 1978 and concluded in March, 1979. Start-up operations were initiated in early March, 1979. The results of the mechanical testing are presented in

sections according to the major functions of the system: SO_2 absorption; absorbent regeneration; solids dewatering; preparation of raw materials; and auxiliary services. In general, all the tanks were checked for leaks and their internal linings were inspected; the pumps and agitators were checked and operated; all the lines were flushed; and all instruments were zeroed and checked for proper operation. Specific tests for each section as well as relevant results are discussed below.

1. Absorber Section

a. Liquid Related

In addition to the general tests, the flow distribution in the absorber trays and the operation of the sprays (tray underside and emergency sprays) was checked. The following corrective actions were required: vibration of the recycle pump motors, initially 4-20 mills, was reduced to below 5 mills by the manufacturer; two butterfly valves in the absorber recirculation loops failed and were replaced by heavy duty butterfly valves; and uneven liquid distribution in the absorber trays was corrected by leveling the trays. Testing of the other items in this section was satisfactory.

b. Gas Related

For this section, mechanical testing consisted of checking the ductwork, the operation of the dampers, checking and balancing the booster fans, checking and operating the gas reheater and checking the instrumentation. Corrective action was required for the booster fans and the dampers.

The fluid drives for the booster fans required a correction in the shaft gap after which the fluid drives were aligned. The control of the gas flow to the absorbers was also modified. A single pressure controller is now used to control the speed of the booster fans. The controller signal is sent to each booster fan through a precalibrated bias. Previously, an independent pressure controller was used for each fan. This change has been incorporated in the control philosophy section (Chapter IV).

The gear boxes for the outlet dampers were replaced and the motors for the damper drives were balanced by the manufacturer. The blades in the inlet dampers had a tendency to drift apart. This problem was caused by defective motor brakes. The possibility of providing additional seals for the dampers to further reduce gas leakage is currently being considered.

2. Reactors

In the reactor section, corrective action was required for the agitators and transfer pumps. An excessive amount of current was being drawn by the agitator motors. The current drawn by the agitators was reduced by trimming the edges off the impellers thereby reducing the impeller size. At the same time, the impeller arms were reinforced to minimize flexing.

Operation of the agitators will be closely checked during start-up operation and any further modifications, if needed, will be included in the Phase III, start-up report.

The large capacity of the transfer pumps caused excessive chatter in the discharge control valves whenever attempting to control low flow rates. To minimize this problem, the pumps were slowed down by changing the pump sheaves.

3. Dewatering Section

a. Thickener and Thickener Hold Tank

During mechanical testing, the thickener overflow weir was adjusted. The thickener rake was checked for rotation and operation and it was leveled. The lining on the rake required some patching. The rake lift mechanism was checked and operated. Some of the rake lift instrumentation was wet and certain parts had to be replaced.

While operating the thickener underflow pumps, an excessive water hammer effect caused the breaking of the lines. The pumps were slowed down to the required range by replacing the pump sheaves.

b. Filters

Mechanical testing of the filters consisted of checking and operating the filter drum and tub agitator drives, the vacuum pumps, and the blow-back fans. The mechanical check-out of the filters did not require any significant corrective action.

4. Raw Materials

a. Lime Slurry Day Tank

The agitator in the lime slurry tank required a modification similar to the one performed on the reactor agitators. The edges of the impeller were trimmed off in order to reduce the current drawn by the agitator motor.

The screens in the suction side of the transfer pump were being plugged by oversized particles. As was previously indicated in Chapter IV, a disintegrator has been installed temporarily until a ball mill is installed in the lime supply system for the Cane Run Station.

b. Soda Ash Make-up

This section of the dual alkali system consists of two major pieces of equipment: the soda ash silo and the solution tank.

For the soda ash silo, the vibrating bin, the dust collector, and the weigh feeder were checked and operated. The weigh feeder was also calibrated.

A bent agitator arm in the solution tank had to be straightened. During loading of the system with soda ash, warm vapor from the solution tank caused crystallization of the soda ash on the feed chute to the solution tank. A blow-back fan was installed to prevent the vapors from entering and plugging the chute.

5. Auxiliary Equipment

In this section of the system, all the sump pumps were checked and operated and all the water supply lines were flushed. The river water used in the system was found to be slimy, coating the walls of rotameters and occasionally hanging the floats. The rotameters are being cleaned frequently while solutions to the problem are being considered.

In general, the individual pieces of equipment performed satisfactorily during the testing period. Very few pieces of equipment presented significant problems; namely, the booster fans, the dampers, the reactor agitators, and pumps.

VI. CAPITAL COSTS FOR THE DUAL ALKALI SYSTEM AT THE CANE RUN STATION

The installation of the dual alkali system at Cane Run Station Unit No. 6 required capital investment in three different facilities: the flue gas desulfurization (FGD), the lime slurry feed system, and a waste processing and disposal system. Each of these facilities have involved independent design and installation efforts, and their costs are reported separately. While construction of the FGD system was essentially completed in February, 1979, a significant amount of work remained on the lime feed and waste processing facilities. Therefore, the costs presented here represent the actual expenditures incurred plus estimates for completion of the system.

Table VI-1 gives a summary of the capital investment for all three facilities. The total projected cost of $20.6 million includes actual expenditures reported through February 28, 1979 and the estimated capital required for completion. Approximately 80% of this total projected capital cost was expended through the end of February. Most of the remaining 20% is related to the waste processing plant.

A further breakdown of capital investment by subsystems is given in Table VI-2. This table also provides information on the capital expenditures incurred through February 28, 1979 as a percent of the estimated total capital required for various cost elements of each facility.

The capital costs in Tables VI-1 and VI-2 have been reported on an "as-incurred plus estimate for completion" basis, and therefore do not represent a constant dollar value of the capital investment. The cash flow records kept by Louisville Gas and Electric have been used to escalate the costs from the time of expenditure to June, 1979. The total capital investment for the dual alkali system (including all three systems) is $22.0 million in June, 1979 dollars.

A. FGD SYSTEM

The total cost of the FGD system is estimated to be $16.5 million. The expenditures incurred as of February 28, 1979 were 92.7% of this value. The remaining 7.3% were the estimated cost for completion. The installation of the FGD system has been completed and start-up operations are in progress. It should be noted that the capital expenditure of $16.5 million includes the cost of the start-up operations. The material costs for the FGD system include all the process equipment, instrumentation, piping and other process related materials supplied by Combustion Equipment Associates and presented in Table VI-3. The material costs also include electrical utilities, service piping, and their associated instrumentation as well as other materials such as foundations, buildings, etc. The breakdown of these additional materials costs is given in Table VI-4.

TABLE VI-1

CAPITAL COSTS FOR THE DUAL ALKALI SYSTEM
AT CANE RUN UNIT NO. 6, LG&E
"AS-INCURRED PLUS ESTIMATE FOR COMPLETION BASIS"

MATERIAL COSTS:	
• FGD System[a]	
- Process Materials	7,392,300
- Additional Materials	
· Utility and Service Piping and Instrumentation	228,900
· Electrical	731,700
· Other	1,740,300
	10,093,200
• Lime Slurry Feed System	
- Process Equipment	617,200
- Electrical, Piping, and Instrumentation	170,800
	788,000
• Solid Waste Disposal System[b]	
- Fly Ash Supply	309,600
- Waste Processing	1,060,900
- Landfill Area Materials[c]	559,000
	1,929,500
TOTAL MATERIAL COSTS	12,810,700
ERECTION COSTS:	
• Direct Labor	3,058,500
• Field Supervision[d]	337,200
• Construction Overhead	2,038,900
TOTAL ERECTION COSTS	5,434,600
ENGINEERING COSTS:[d]	
• System Supplier's Engineering	1,162,700
• Owner's Consultant Engineering	985,000
TOTAL ENGINEERING COSTS	2,147,700
SPARE PARTS:	203,900
TOTAL CAPITAL INVESTMENT	20,596,900

[a]See Tables VI-3 and VI-4 for breakdown of costs.
[b]Battery limits: cake discharge from filters.
[c]Includes earth moving equipment, dikes, culverts, and bridges.
[d]Owner's engineering is included in field supervision.

TABLE VI-2

CAPITAL COST BREAKDOWN BY SUB-SYSTEM[a]
LOUISVILLE GAS AND ELECTRIC
CANE RUN UNIT NO. 6

Sub-System	Material Costs[b]	Erection Costs			Engineering Costs		Total
		Direct Labor	Field Superv. & Engineering	Construction Overhead	Suppl.	LG&E Consult.	
FGD	10,256,200	2,400,900	281,800	1,376,400	1,162,700	985,000	16,463,000
	(91%)	(92%)	(98%)	(92%)	(100%)	(100%)	(93%)
Lime Slurry	800,000	205,500	16,400	194,300		c	1,216,200
	(69%)	(72%)	(70%)	(71%)			(70%)
Waste Disposal[d]	1,958,400	452,100	39,000	468,200		c	2,917,700
	(0%)	(0%)	(0%)	(0%)			(0%)
TOTAL	13,014,600	3,058,500	337,200	2,038,900	1,162,700	985,000	20,596,000

[a] Numbers in parentheses represent expenditures incurred as of February 28, 1979 as percent of the estimated total costs.

[b] Includes spare parts at 1% of total costs.

[c] Included in Field Supervision and Engineering.

[d] The estimates for waste disposal system are based on a contract awarded to IUCS to provide LG&E with the waste disposal facilities and estimated costs for the erection of the system.

TABLE VI-3

MATERIAL COSTS FOR THE FGD SYSTEM[a]

- PROCESS MATERIALS -

Process Equipment	
- Absorbers	901,510
- Tanks	609,605
- Thickener Shell and Mechanism	340,191
- Vacuum Filters	971,298
- Reheaters	167,681
- Fans and Fluid Drives	570,249
- Pumps and Motors	835,316
- Agitators	95,905
- Weigh Feeder	49,911
- Lining	424,739
	4,966,405
Process Instrumentation	439,100
Process Piping, Insulation, and Heat Tracing	634,568
Other Process Materials	
- Ductwork and Dampers	1,028,702
- Structural Steel	323,543
	1,352,245
	7,392,318

[a]As-incurred plus estimate for completion basis.

TABLE VI-4

MATERIAL COSTS FOR THE FGD SYSTEM[a]

- ADDITIONAL MATERIALS -

Utility and Service Piping and Instrumentation	
- Instrumentation	73,300
- Piping, Insulation, and Heat Tracing	86,200
- Miscellaneous	69,400
	228,900
Electrical	
- Electrical Auxiliaries (4160, 460, 220/110 volt)	487,900
- Station Grounding and Ducts	38,100
- Station Control (wiring to instruments)	205,700
	731,700
Other	
- Foundations[b]	964,700
- Buildings	220,200
- Land Improvements, Roads, and Storm Runoff Containment	555,400
	1,740,300

[a]As-incurred plus estimate for completion basis.

[b]Includes lagging and insulation for fans.

B. LIME SLURRY FEED

The lime slurry system encompasses all the facilities required to unload, store, and process the carbide lime slurry being barged from a nearby acetylene plant. Part of these facilities existed prior to the installation of the dual alkali system, but had to be expanded and modified--in particular, the piping from the dock, the pumping, and the storage facilities. This portion of the system has been completed. However, since the carbide lime contains some oversized material, a grinding mill consisting of a hydroclone and wet ball mill are being installed. It is estimated that this system will be completed in June, 1979.

The lime slurry facilities are used by all three FGD systems on Unit Nos. 4, 5, and 6 at the Cane Run Station. The capital costs have been apportioned to the dual alkali plant according to the relative operating capacity of the units. (Unit No. 6 capacity represents roughly 3/7 of the total capacity of all three boilers.)

The total cost of the lime slurry system attributed to the dual alkali system is estimated to be $1.2 million. The expenditures as of February 28, 1979 amounted to 70% of this value. The remaining 30% is the estimated cost of the grinding mill.

C. WASTE DISPOSAL

The cost of the waste processing and disposal system is estimated at $2.9 million. As of February 28, 1979, no expenditures were incurred for this system. The estimated cost, therefore, is based on the contract awarded to I.U. Conversions Systems (IUCS) for providing LG&E with the equipment needed for this system, and on estimated costs for the erection of the system. The waste disposal system involves stabilization of the FGD waste with fly ash and lime and dry landfill of the stabilized material in a location adjacent to the plant. The material costs include: the discharge conveyors to transport the FGD waste from the filter to the pug mill; the fly ash and lime supply; and pug mill for mixing the filter cake, fly ash and lime; the stacker conveyors; and disposal equipment.

VII. GLOSSARY

Active Sodium - Sodium associated with anions involved in SO_2 absorption reactions and includes sulfite, bisulfite, hydroxide and carbonate/bicarbonate. Total active sodium concentration is calculated as follows:

$$[Na^+]_{active} = 2 \times ([Na_2SO_3] + [Na_2CO_3]) + [NaHSO_3] + [NaOH] + [NaHCO_3]$$

Available Alkali - The percentage of the calcium hydroxide in the raw hydrated lime, or in the insoluble solids in the carbide lime slurry.

Calcium Utilization - The percentage of the calcium in the lime or limestone which is present in the solid product as a calcium-sulfur salt. Calcium utilization is defined as:

$$\text{Calcium Utilization} = \frac{\text{mols } (CaSO_3 + CaSO_4) \text{ generated}}{\text{mol Ca fed}} \times 100\%$$

Concentrated Dual Alkali Modes - Modes of operation of the dual alkali process in which regeneration reactions produce solids containing $CaSO_3 \cdot 1/2\ H_2O$ or a mixed crystal containing calcium sulfite and calcium sulfate hemihydrates, but not containing gypsum. Active sodium concentrations are usually higher than 0.15M Na^+ in concentrated mode solutions.

Dilute Dual Alkali Modes - Modes of operation of the dual alkali process in which regeneration reactions produce solids containing gypsum ($CaSO_4 \cdot 2\ H_2O$). Active sodium concentrations are usually lower than 0.15M Na^+ in dilute mode solutions.

Sulfate Formation - The oxidation of sulfur (IV)-sulfite and bisulfate. The level of sulfate formation relative to SO_2 absorption is given by:

$$\text{Sulfate Formation} = \frac{\text{mols S (IV) oxidized}}{\text{mol } SO_2 \text{ removed}} \times 100\%$$

Sulfate Precipitation - The formation of $CaSO_4 \cdot XH_2O$ insoluble solids. The level of sulfate precipitation in the overall scheme is given by the ratio of calcium sulfate to the total calcium-sulfur salts produced:

$$\text{Sulfate Precipitation} = \frac{\text{mols } CaSO_4 \text{ produced}}{\text{mols } (CaSO_3 + CaSO_4) \text{ produced}}$$

APPENDIX

EQUIPMENT DETAILS

TABLE A-1

EQUIPMENT LIST

Equipment	Description	No. Required
1. Agitators (w/motors)	Reactant Feed Tank	1
	Primary Reactor Tanks	2
	Secondary Reactor Tanks	2
	Soda Ash Solution Tank	1
	Vacuum Filters	3
2. Dampers	Booster Fan Inlets	2
	By-Pass	1
	Absorber Outlets	2
	Reheater Fans	2
3. Ductwork	Take-Off Connecting Ducts	2
	Booster Fan Inlet Ducts	2
	Booster Fan Outlet Ducts	2
	Absorber Outlet Ducts	2
	Duct/By-Pass Transition	1
4. Expansion Joints	Tie-Ins	2
	Booster Fan Inlets	2
	Booster Fan Pant Legs	4
	Booster Fan Outlets	2
	Booster Fan Outlet Ducts	2
	Absorber Outlet PC#1	2
	Absorber Outlet PC#2	2
	Breeching	1
	By-Pass	2

TABLE A-1
(Continued)

Equipment	Description	No. Required
5. Fans	Booster Fans, Drives, & Motor	2
	Reheater Fans & Motors	2
	Filter Blow Back Fans & Motors	3
	Soda Ash Chute Fan & Motor	1
6. Heaters	Reheater	2
7. Pumps & Feeders	Lime Reactant Feed Pump & Motor	2
	Reactor Transfer Pump & Motor	2
	Thickener Underflow Pump & Motor	2
	Soda Ash Solution Pump & Motor	2
	Thickener Hold Tank Transfer Pump & Motor	2
	Absorber Recycle Pump & Motor	4
	Vacuum Pumps & Motor	3
	Filtrate Sump Pumps & Motor	2
	Soda Ash Sump Pumps	2
	Silencer Overflow Sump Pumps & Motors	2
	Soda Ash Weigh Feeder & Motor	1
8. Tanks	Reactant Feed Tank	1
	Primary Reactor Tank	2
	Secondary Reactor Tank	2
	Soda Ash Silo	1
	Thickener Hold Tank	1
	Soda Ash Solution Tank	1
	Filtrate Receivers	3
	Thickener	1

TABLE A-1
(Continued)

	Equipment	Description	No. Required
10.	Thickener Mechanism	Thickener Mechanism & Motor	1
		Thickener Lift Rake Motor	1
11.	Vacuum Filter	Vacuum Filters & Drives	3
12.	Vessels	Absorbers	2

TABLE A-2

MATERIALS OF CONSTRUCTION

1. Agitators

   (a) Reactant feed tank--carbon steel.

   (b) Primary reactors--shaft: carbon steel, rubber lined; hub and blades: 317L s.s.

   (c) Secondary reactors--shaft: carbon steel, rubber lined; hub and blades: 317L s.s.

   (d) Soda ash solution tank--carbon steel, rubber lined.

2. Dampers

   (a) Booster fan inlet--A-283 carbon steel.

   (b) Bypass and absorber outlet--317L s.s.

3. Ductwork

   (a) Ductwork carrying hot flue gas to the scrubber inlet--carbon steel.

   (b) Ductwork carrying saturated flue gas from the scrubber to reheater--carbon steel, coated internally with a flake reinforced polyester lining.

   (c) Ductwork carrying reheated flue gas from the reheater to stack--317L s.s.

   (d) Bypass/transition duct--carbon steel.

4. Expansion Joints

   (a) Expansion joints on inlet side of absorbers--viton.

   (b) Expansion joints on outlet side of absorbers--chlorobutyl for wet gas and viton for dry gas (after reheat).

5. Fans

   (a) Housing--A441.

   (b) Blades--A441, with ware plates constructed from A441 material.

TABLE A-2
(Continued)

6. Pumps

   (a) Housing--rubber lined.

   (b) Impeller--rubber lined.

7. Tanks

   (a) Thickener hold tank--carbon steel, flake reinforced polyester lining.

   (b) Primary reactors--316L s.s.

   (c) Secondary reactors--carbon steel, flake reinforced polyester lining, rubber pad on bottom.

   (d) Reactant feed tank--carbon steel.

   (e) Soda ash solution tank--carbon steel.

8. Soda Ash Silo

   (a) Carbon steel.

9. Weigh Feeders

   (a) Frame--mild steel.

   (b) Internals--304 s.s.

10. Thickener

   (a) Thickener shell--carbon steel, bottom concrete with interior, flake reinforced lining.

   (b) Rake, shaft, and centerwell--carbon steel, rubber lined.

11. Vacuum Filter

   (a) Filter drum--317 ELC s.s.

   (b) Agitator--317 ELC s.s.

   (c) Filtrate receiver--FRP.

TABLE A-2
(Continued)

12. Absorber

(a) Absorber shell--carbon steel, coated internally with flake reinforced polyester lining.

(b) Absorber trays--317 s.s.

(c) Demister--noryl.

13. Piping

(a) All process piping--FRP.

(b) Piping for make-up water and service water and all other piping not subject to corrosion--carbon steel.

TABLE A-3

AGITATORS

	Reactant Feed Tank	Primary Reactors	Secondary Reactors	Soda Ash Solution Tank
Number Required:	1	2	2	1
Impeller Type	turbine	turbine	turbine	propeller
Impeller dia.	89"	67"	95"	12"
RPM	30	37	30	350
Shaft dia.	4.5"	3.5"	5"	1.5"
H.P.	20	7.5	25	1.5
Material of Construction:				
Shaft	C.S.	R.C.C.S.	R.C.C.S.	R.C.C.S.
Blades	C.S.	317L S.S.	317L S.S.	R.C.C.S.

Data are given per agitator
C.S. - Carbon Steel
R.C.C.S. - Rubber Covered Carbon Steel
S.S. - Stainless Steel

TABLE A-4

DAMPERS

	Booster Fan Inlet	Bypass	Absorber Outlet
Number Required:	2	1	2
Design flow rate			
- ACFM	533,000	1,065,000	487,000
- °F	350	350	200
Size	135-1/4" x 138-1/2"	162" x 240"	156" dia.
Position of duct	Horizontal	Horizontal	Horizontal
Type	Guillotine	Multi-louver	Guillotine
Entry	Bottom	--	Top
Normal position	Open	Closed	Open
Material of construction	A-36 steel	316L S.S.	316L S.S.
Max. gas leakage, % of design flow rate	1	1	1
Paint external members	Zinc chromate	Zinc chromate	Zinc chromate

TABLE A-5

DUCTWORK

Equipment	Number Required	Dimensions
Take off connecting duct	2	Inlet 11'8" x 11'6" Outlet to booster fan inlet duct 11'3" x 11'6" Outlet to bypass transition 11'1" x 12'1" Overall dimensions 11'8" x 11'6" x 16'10"
Booster fan inlet duct	2	Inlet 11'3" x 11'6" Outlets 16'8" x 3'2" (two) Overall dimensions 11'6" x 28' x 16'6"
Booster fan outlet duct	2	Inlet duct 10'7" d Outlet duct 13'9" d Overall length 12'
Absorber outlet duct	2	Inlet 13' d Outlet 13' d Overall length 80'
Duct bypass/transition	1	Inlets from take off connecting duct 11'1" x 12'1" (two) Inlets from absorber outlet duct 13' d (two) Outlet 28' x 13'6" Overall dimensions 30'3" x 24'6" x 19'10"

Note: See Table A-2 for materials of construction

TABLE A-6

EXPANSION JOINTS

Service	Number Required	Size	Material
Tie-In (DJ 101/201)	2	11'10" x 11'8-12"	Viton
Booster Fan Inlet (DJ 102/202)	2	11'5-1/4" x 11'8-1/2"	Viton
Booster Fan Pant Leg (DJ 103A/203A) (DJ 103B/203B)	4 2	3'3-7/8" x 16'10-1/4"	Vitron
Booster fan outlet (DJ 104/204)	2	10'9-1/4" x 8'9-1/2"	Viton
Absorber inlet (DJ 105/205)	2	13'11-5/8" dia.	Viton
Absorber outlet 1 (DJ 106/206)	2	13'2" dia.	Chlorobutyl
Absorber Outlet 2 (DJ 107/207)	2	13'2" dia.	Viton
Breeching (DJ 108)	1	13'7-7/8" x 28'1-7/8"	Viton
Bypass (DJ 109)	2	11'10" x 11'8-1/2"	Viton
Reheater duct (DJ 110/210) (DJ 111/211)	4	6'9-1/4"	Special Hi-Temp. Material

All the expansion joints except absorber outlet 1 and reheater duct expansion joints are designed for 400°F with excursions to 600°F for 5 minutes, 4 times a year. All the expansion joints are 9" wide, except DJ 107/207 - 16" wide and DJ 110/210, 111/211 - 11-1/4" wide.

TABLE A-7

BOOSTER FANS

Number required	2
Flue gas volume	533,000 acfm
Flue gas temperature	300°F
Inlet static pressure	±2 inch WG
Outlet pressure at design flow rate	10.5 inch WG
Gas density	0.0526 lb/cu ft
Inlet dust loading	0.0537 gr/cu ft
Maximum vibration amplitude	2.2 mils at 720 rpm
Type	Centrifugal forced draft
Fan blade design	Backward inclined airfoil with wear plates
Materials of construction	Carbon steel
Drive	Fluid drive
Motor HP	1,300
Volts	4,000
Motor rpm	720

TABLE A-8

REHEATERS

Operating Conditions	
Wet flue gas flow rate	436,500 acfm
Temperature	126°F
Pressure	+2 inch WG
H_2O Vapor	2,475 lbs/min
SO_2	11.25 lbs/min
	200 ppm dry basis
Particulates	2.48 lbs/min
Temperature after reheat	176°F
Heater Requirements	
For flue gas	25,632,000 Btu/hr
Radiation loss	1,282,000 Btu/hr
Heater outlet temperature	800°F max.
Turndown	To 20% of the capacity
Heater outlet pressure	+7 inch WG
Fuel Data	
Fuel type	Number 2 fuel oil
Oil flow rate	171 gal/hr
Air Inlet Temperature	
Winter	0°F
Summer	100°F
No. of Reheaters Required	2

TABLE A-9

PUMPS

	Absorber Recycle Pump	Reactor Pump	Thickener Hold Tank Pump	Thickener Underflow Pump	Reactant Feed Pump	Soda Ash Solution Pump
Number Required						
Operating	2	2	1	1	1	1
Spare	2	-	1	1	1	1
Capacity						
gpm	4,600	1,965	4,185	665	340	140
head, ft	130	85	105	115	115	80
Speed	Variable	Variable	Variable	Variable	Variable	Variable
Material of Construction	RLCI	RLCI	RLCI	RLCI	NI-Hard	RLCI
Packing	Yes	Yes	Yes	Yes	Yes	Yes
Drive	V belt	V belt	V belt	V belt	V belt	V belt
Motor Mounting	Overhead	Overhead	Overhead	Overhead	Overhead	Overhead
Voltage, volts	4,000	460	4,000	460	460	460
Drip proof	Yes	Yes	Yes	Yes	Yes	Yes
BHP/IHP	215/250	62/75	157/200	33/40	25/25	6/10
Service Factor	1.15	1.15	1.15	1.15	1.15	1.15
Overall Size[a]	12 x 10 x 25	10 x 8 x 21	12 x 10 x 25	5 x 5 x 14	3 x 1.5 x 16	1 x 1.5 x 6

RLCI - Rubber lined cast iron
All pumps are centrifugal pumps
[a]All dimensions are in feet

TABLE A-10

TANKS

Process Data	Primary Reaction Tank	Secondary Reaction Tank	Thickener Hold Tank	Reactant Feed Tank	Soda Ash Solution Tank
Liquor specific gravity	1.1	1.1	1.1	1.2	1.1
pH range	5-11	11-12.5	12	12	12
Chlorides, ppm	12,000	12,000	12,000	---	12,000
Operating pressure, in wg	Liquid head	Liquid head	Liquid head	Liquid head	Liquid head
Design pressure, in wg	Liquid head	Liquid head	Liquid head	Liquid head	Liquid head
Operating temperature, °F	126	126	110	70	110
Design temperature, °F	---	---	---	100	---
Specified data					
Minimum thickness, inches	3/16	1/4	1/4	1/4	1/4
Seismic zone	1	1	1	1	1
Code	API650	API650	API650	API650	API650
Tank shape	Cylindrical	Cylindrical	Cylindrical	Cylindrical	Cylindrical
Dimensions, dia. x height	11' x 14'	20' x 33'	36' x 23'	24'6" x 24'6"	6' x 8'
Baffles	4	4	None	4	4
Agitator	Yes	Yes	No	Yes	Yes
Materials					
Shell and head	316L S.S.	A283 (C.S.)	A283 (C.S.)	A283 (C.S.)	A283 (C.S.)
Internal structure	316L S.S.	C.S.	C.S.	C.S.	C.S.
Nozzles necks/flanges	316L S.S.	C.S. & 316L S.S.	C.S. & 316L S.S.	C.S.	C.S. & 316L S.S.
Lining	None	Glass reinforced polyester + 3/8" thick rubber pad on bottom	Glass reinforced polyester	---	Glass reinforced polyester
Paint	Zinc chromate primer	Zinc chromate primer	Zinc chromate primer	Zinc chromate primer	Zinc chromate primer
Gaskets	Neoprene	Neoprene	Neoprene	Neoprene	Neoprene
Erection weight, lbs	7,500	36,000	39,000	39,000	2,500
Operating weight, lbs	97,500	715,000	1,183,000	785,000	17,500

C.S. - carbon steel; S.S. - stainless steel; all tanks are closed at top except thickener hold tank

TABLE A-11

SODA ASH SILO

Process Data	
Specific gravity	65 lbs/ft^3
Soda ash type	dense
Specified Data	
Minimum thickness	1/4"
Wind load at 30'	30 PSF
Seismic zone	1
Code	API650
Tank shape	Cylindrical plus conical bottom
Size	12' diameter x 34'6" height-cylinder 6' bottom diameter x 5'6" height-cone
Thickness	3/8" thick bottom of cylinder 5/16" thick middle of cylinder 1/4" thick top of cylinder 1/4" thick cone
Attachments	Cone vibrator, baghouse at top, and piping at top for pneumatic feed system
Materials	
Shell and head	A283 (C.S.)
Internals	C.S.
Nozzles necks/flanges	C.S.
Paint	Zinc chromate primer
Gaskets	Neoprene
Erection weight	22,000 lbs
Operating weight	300,000 lbs

TABLE A-12

THICKENER

Design stream conditions	
Reactor bleed	3,570 gpm
Filter overflow	148 gpm
Filtrate	500 gpm
Soda ash solution	50 gpm
Thickener recycle	155 gpm
Thickener underflow	605 gpm
Operating temperature	110°F
Seismic zone	1
Thickener type	Flat bottom
Diameter	125'
Height	23'
Feed well	
Diameter	12'
Height	11'
Number of arms on rake	2
Cone scraper	Yes
Overflow weir plate	Notched
Access walk way	On one side of the superstructure
Rake drive motor	5 HP
Rake lifting device motor	3 HP
Materials	
Rake	Rubber covered carbon steel
Feed well	Rubber covered carbon steel
Shell	Carbon steel
Bottom	Concrete
Weir plate	Polypropylene
Shell and bottom	Lined with glass reinforced polyester
Paint	Zinc chromate primer

TABLE A-13

VACUUM FILTER

Process data	
Net slurry feed	Design 3,147 lbs/min - 302 gpm Maximum 4,533 lbs/min - 448 gpm
Specific gravity	1.25
Temperature	140°F maximum
pH	11-12.5
Wash water	2,108 lbs/min design
Filter requirements	
% Solids in cake	55% minimum 63% average
Total flow	1,246 lbs/min design 1,584 lbs/min maximum
Filtrate	
Flow rate	299-750 gpm
pH	10.5-12.3
Specific gravity	1.064
Cake wash rate	91 gpm normal 300 gpm maximum
Number of filters required	3 including spare
Size	8' diameter 16' face
Cake discharge mechanism	Blower assisted scraper blade
Liquid level control in filter tub	Adjustable overflow weir
Cake wash assembly	Drip wash nozzles with drag net
Filter drum speed	0.3-3.0 rpm
Agitator type	Counter weighted rocker arm
Filtrate receiver	54" diameter x 54" height
Motor	
Vacuum pump	100 HP
Drum drive	5 HP
Filter blow back pump	2 HP
Agitator drive	1.5 HP
Materials	
Filter drum	317 ELC stainless steel
Filter agitator	317 ELC stainless steel
Filter scraper	317 stainless steel with rubber tip
Filtrate receiver	FRP
Drainage grid	Polypropylene
Filter media	Polypropylene

TABLE A-14

ABSORBER

Process data	
Specific gravity	1.2
pH range	5-12
Chlorides, ppm	12,000
Operating pressure, inch WG	+11.5
Design pressure, inch WG	+12.5
Operating temperature, °F	125
Design temperature, °F	350
Upset conditions	
Temperature, °F	600
Time, minutes	5
Specified data	
Corrosion allowance	none
Wind load at 30', PSF	30
Seismic zone	1
Code	API650
Tank shape	Cylindrical shell with conical head
Size	32' diameter x 45' height shell 10'6" height x 13' top diameter cone
Thickness, inch	3/8
Internals	Sprays + 2 trays + chevron demister
Materials	
Shell and head	A283 (carbon steel)
Trays/supports	317L S.S.
Spray nozzles	316 S.S. & 317 S.S.
Internal piping/supports	FRP
Mist eliminator/supports	Noryl
Nozzle necks/flanges	C.S. and S.S.
Internal fasteners	C.S.
Gaskets	Neoprene
External paint	Zinc chromate
Lining	Glass reinforced polyester
Erection	
Erection weight, lbs	108,000
Operating weight, lbs	475,000